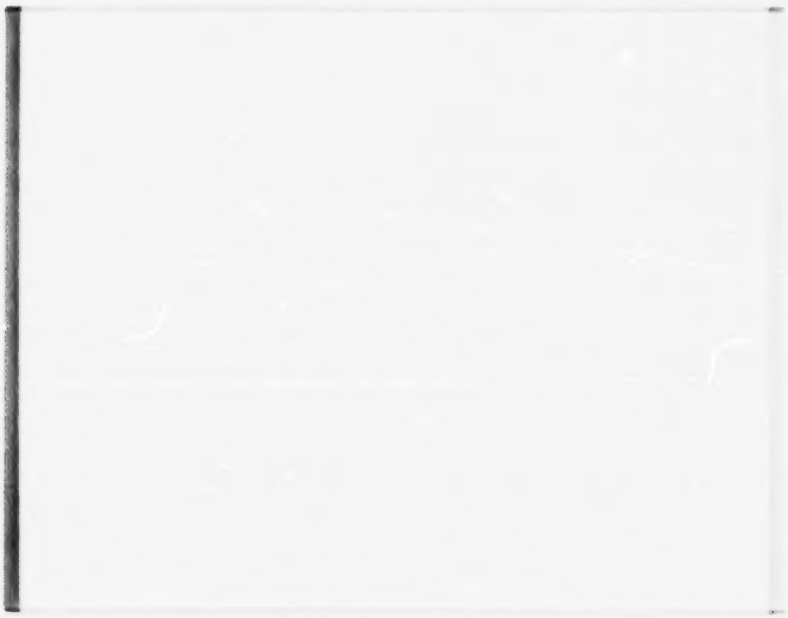
MINERALS

ECONOMIC & REGIONAL DEVELOPMENT AGREEMENT



Canada

Manitoba

MG-13097A

BENCHSCALE EVALUATION OF THE APPLICATION OF THE GREAT CENTRAL MINES HYDROMETALLURGICAL COPPER PROCESS TO HUDSON BAY MINING AND SMELTING COPPER CONCENTRATES

ENGINEERING AND ECONOMIC STUDY

1987 JUNE

MICROMEDIA

BENCHSCALE EVALUATION OF THE APPLICATION OF THE GREAT CENTRAL MINES HYDROMETALLURGICAL COPPER PROCESS TO HUDSON BAY MINING AND SMELTING COPPER CONCENTRATES

ENGINEERING AND ECONOMIC STUDY

by

B.D. Krysa - HBM&S
H.H. Brown - HBM&S
W.J.S. Craigen - CANMET, EMR

1987 JUNE

Work on the project was conducted under the auspices of the Canada Centre for Mineral and Energy Technology, Energy, Mines and Resources Canada, under the Canada/Manitoba Mineral Development Agreement.

MICROMEDIA

DSS file number: 23SQ.23440-4-9209

DSS contract serial number: OSQ84-00310

68 pp

TABLE OF CONTENTS

		PAGE NO.
1.	EXECUTIVE SUMMARY	1
2.	INTRODUCTION	3
3.	THE GCM HYDROCOPPER PROCESS	
3.1	BACKGROUND	4
3.2	PROCESS DESCRIPTION	6
4.	ENGINEERING STUDY - DESCRIPTION AND SCOPE	8
5.	PLANT LAYOUT AND LOCATION	11
6.	CAPITAL COSTS	13
7.	OPERATING COSTS	13
8.	FINANCIAL EVALUATION	14
9.	CONCLUSIONS	16
10.	RECOMMENDATIONS	17
11.	REFERENCES	19
	APPENDIX A PROCESS DESIGN CRITERIA	21
TABLE I	CONCENTRATE ANALYSIS AND MINERALOGY	22
TABLE II	CHEMICAL REACTIONS	23
TABLE III	EQUIPMENT DESIGN CRITERIA	24
	APPENDIX B MASS BALANCE	29
FIGURE 4	FLOWSHEET STREAM IDENTIFICATION	30
TABLE IV	LIST OF UNIT OPERATIONS	31
TABLE V	LIST OF PROCESS FLOWS	32
FIGURE 5	MAJOR STREAM FLOWS	33
TABLE VI	DETAILED STREAM FLOWS	34
	APPENDIX C ENERGY BALANCE	44
TABLE VII	GCM PROCESS ENERGY REQUIREMENTS	45
	APPENDIX D CAPITAL COSTS	46
TABLE VIII	CAPITAL COST SUMMARY	47
TABLE IX	CAPITAL COST DETAILED ESTIMATE	48
TABLE X	CAPITAL COST COMPARISON	49
TABLE XI	MECHANICAL EQUIPMENT LIST	50
TABLE XII	PROCESS FLOW PUMP LIST	54
	APPENDIX E OPERATING COSTS	56
TABLE XIII	OPERATING COSTS	57
TABLE XIV	BASIS FOR OPERATING COSTS	58

APPENDIX F FINANCIAL EVALUATION 59
TABLE XV FINANCIAL ANALYSIS SUMMARY 60
TABLE XVI FINANCIAL ANALYSIS CRITERIA - METAL DISTRIBUTION 61
TABLE XVII FINANCIAL ANALYSIS CRITERIA - INCREMENTAL OUTPUTS AND REVENUES 63
TABLE XVIII FINANCIAL ANALYSIS CRITERIA - BASIS FOR REVENUES 64
TABLE XIX FINANCIAL ANALYSIS CRITERIA - INCREMENTAL CHANGES IN OPERATING COSTS 66

1. EXECUTIVE SUMMARY

Based on laboratory scale testwork, an engineering and economic evaluation has been carried out on the application of the Grent Central Mines (GCM) Hydrometallurgical copper process for the treatment of HBM&S copper concentrates at Flin Flon, Manitoba. The benefits of implementing the GCM process are measured from a "base case", which assumes that the Zinc Pressure Leach (ZPL) process has already been implemented at Flin Flon, for the processing of zinc concentrates.

The assessment indicates that the GCM process benefits of $7.3 million/y are insufficient to justify the large capital expenditure ($157 million) required to build the plant. Two major factors have a dramatic effect on GCM process benefits for HBM&S. These are copper refining costs, where savings of about $13 million/y are indicated, and gold revenues, where losses of about $14 million/y are indicated. If the latter can be substantially reduced by developing an on-site facility for gold recovery, overall benefits for the GCM process could increase up to a maximum of about $20.9 million/y. This would provide a DCF ROI of about 10% over a 15 year period. This figure is substantially lower than previous (1984) estimates (1), primarily because savings in copper refining costs are lower, and savings in operating costs for the GCM process over the base case have not materialized. The latter is due mainly to improved copper smelter productivity, loss of fuming plant credits to the ZPL process and higher than originally estimated GCM process operating costs.

Even if the gold revenue problem is successfully resolved, the GCM process cannot be commercially implemented until it has been shown that cathode quality copper can be consistently produced and the savings in copper refining costs verified. To accomplish this, continuous pilot scale testwork on the GCM process is required, at a cost that will probably be in the order of + $5 million. In addition to copper purity, other factors that have a significant effect on the GCM process operating costs and benefits should also be investigated. These are:

- The relative viability of producing a chloride-free zinc product by H_2SO_4 addition, either in the preleach or during oxyhydrolysis.

- Better methods to separate sulphur from the leach residue, to maximize sulphur recovery and minimize the volume of residue which has to be treated for gold recovery.

- A method to improve the purity and saleability of the sulphur by-product.

On-site methods for recovering the gold from hydrometallurgical (GCM; ZPL) residues are already being investigated. While the economic viability of the project is in doubt, if the results of the above studies prove positive, it could be recommended that continuous pilot scale testing of the GCM process be considered, based on its potential to economically produce copper while eliminating SO_2 emissions.

BENCHSCALE EVALUATION OF THE G.C.M. HYDROMETALLURGICAL COPPER PROCESS ENGINEERING STUDY

2. INTRODUCTION

In 1983, Energy, Mines and Resources (EMR) Canada, in co-operation with industry, Environmental Canada and the Department of Regional Industrial Expansion (DRIE), undertook a comprehensive study of Canada's nonferrous metals industry. The objectives of the study were to review the productivity and international competitiveness of the industry, to analyze the impact of more stringent environmental controls on long term viability, and to recommend modernization/pollution abatement strategies and technology development initiatives consistent with future economic and environmental goals. Results of this study were documented in two EMR special reports; one on nickel and copper, issued in May 1984 (1), and one on lead and zinc, issued in May 1985 (2). More detailed, "Technological Profile" reports, assessing the technological alternatives and research and development requirements were also issued by CANMET, on a site specific basis for selected nickel and copper smelters (3, 4, 5, 6).

One of the nonferrous metal producers investigated in the aforementioned study was Hudson Bay Mining and Smelting (HBM&S). HBM&S operates a copper-zinc smelter complex at Flin Flon, Manitoba. At present all the SO_2 produced at the smelter complex (258,000 tonnes/y) is emitted to atmosphere. In order to reduce SO_2 emissions and improve productivity a number of selected alternatives were evaluated to replace or modify the existing smelter flowsheet. Because of the smelter's relatively remote inland location and long distance from sulphuric acid markets, hydrometallurgical alternatives which produce elemental sulphur were given high priority in the selection process.

SO_2 or S

In the preliminary evaluation carried out by CANMET (6), the Sherritt zinc pressure leach process (7) and the Great Central Mines (GCM) hydrometallurgical copper process (8, 9) were identified as potentially the most attractive alternatives for replacing the zinc and copper smelters at Flin Flon. This report provides a more detailed engineering and economic evaluation of the GCM process, based on laboratory scale testwork carried out on HBM&S copper concentrates.

3. THE GCM HYDROCOPPER PROCESS

3.1 BACKGROUND

The GCM hydrocopper process was developed through the combined efforts of Great Central Mines Ltd. and Bacon, Donaldson and Associates Ltd. (BDA), both of Vancouver, B.C. The process is patented and wholly owned by Great Central Mines Ltd. The process technology was developed on a laboratory scale at the Bacon, Donaldson laboratories over the period 1976 to 1982.

An evaluation of the process was carried out by CANMET in June 1980 (10), and subsequently in November 1982 (11), and October 1984 (8), after certain recommended improvements were incorporated into the process flowsheet and the feasibility of recovering silver and copper separately was confirmed.

The GCM hydrocopper process (8, 9) is a chloride based process capable of recovering high purity copper, and silver and sulphur by-products, from chalcopyrite concentrates. The recovery of a metallic iron by-product also appears feasible if economics so warrant.

All of the original testwork on the process was carried out using high grade British Columbia porphyry type copper concentrates, which have relatively low levels of impurities (As, Sb, Pb, Zn, etc.). To evaluate the process for application to relatively impure, low grade HBM&S copper concentrates, Great Central Mines was requested to prepare a proposal in co-operation with Bacon, Donaldson and Associates to carry out benchscale testwork under contract to HBM&S/CANMET, with funding for the project provided through the Canada/Manitoba Mineral Development Agreement (MDA). Testwork was initiated in February 1985, and a final detailed confidential report (12), in three volumes, was submitted by BDA in June 1986. The report indicated that application of the GCM process to HBM&S Copper Concentrate was technically feasible provided the process was modified to incorporate a number of additional unit operations, which were necessary to minimize the effect of the impurities present. A conceptual flowsheet, design criteria and mass balances for a plant processing 1000 tonnes per day of concentrate were provided by BDA, and approximate capital and operating costs were estimated, based primarily on an update of previous cost data. To improve the accuracy of these estimates

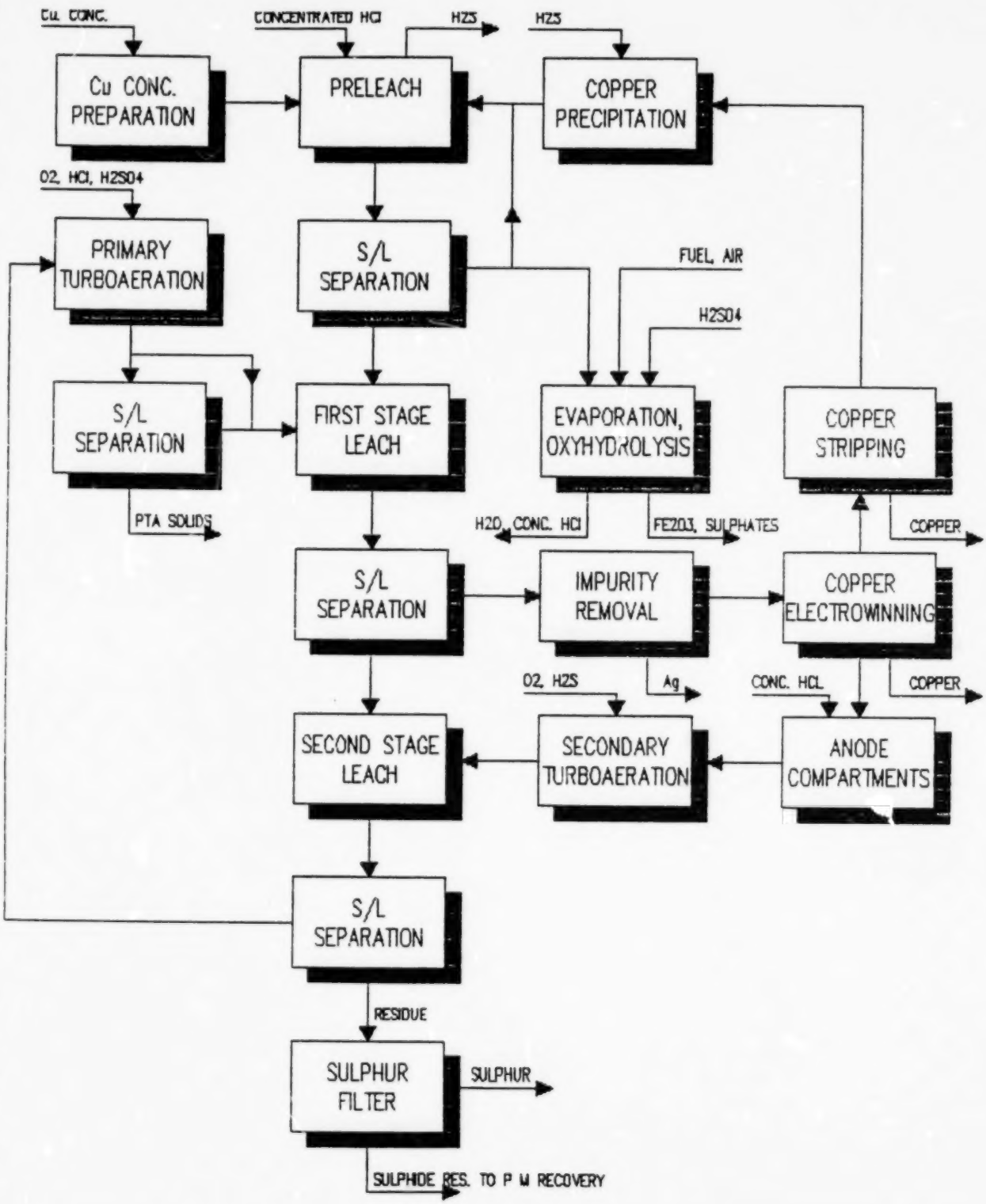


Figure 1 GCM / HBMS Flowsheet

and provide a complete financial analysis for application of the GCM process at Flin Flon, HBM&S undertook a detailed engineering study based on the BDA design criteria and material balances. This study, which is the subject of this report, was carried out under contract to CANMET to meet the overall objectives of this Canada/Manitoba MDA project.

3.2 PROCESS DESCRIPTION

For application to HBM&S Copper Concentrate, the GCM process was modified to include the following additional unit operations:

a) a preleach step to remove impurities from the concentrate;

b) a copper sulphide precipitation step (using H_2S generated in the preleach), to remove copper from the iron bleed stream from electrolysis; and

c) a primary turboaeration step to precipitate iron and impurities (As, Sb, etc.) prior to electrowinning.

A conceptual flowsheet is shown in Figure 1. In the modified process, the concentrate is ground to 80% minus 28 micron and preleached with HCl to remove pyrrhotite, zinc and lead. After solid/liquid separation, the leach solution, which is copper free, is sent directly to the evaporator/oxyhydrolysis circuit where water is removed to maintain the water balance, iron is converted to hematite, non-ferrous metals are converted to sulphates, and HCl is regenerated for recycle to the process. The solids from the preleach are leached with acidic ferric/cupric chloride at atmospheric pressure to extract at least 99% of the copper and 93% of the silver, and to convert reactive sulphides to elemental sulphur. A two-stage (reduction/oxidation) leach with inter-stage thickening is used to ensure high metal extractions, and that the pregnant solution going to the electrowinning circuit is in the cuprous/ferrous state.

After silver removal, the metal chloride solution is electrolyzed in diaphragm cells to produce a granular copper product and regenerate a portion (approximately 30%) of the ferric/cupric chloride lixiviant in the anolyte compartment. Regeneration of the remainder of the lixiviant is carried out with oxygen in two

turboaerators. The secondary turboaerator (STA) is operated under acidic conditions to prevent iron precipitation during regeneration of ferric chloride, which is then recycled to the second (oxidation) stage of leaching. The primary turboaerator (PTA) is operated under more neutral conditions to promote precipitation of about 25% of the iron. About 1/3 of this precipitate is recycled, along with the regenerated lixiviant, to the first (reduction) stage of leaching to absorb deleterious impurities from solution prior to electrowinning. The remaining 2/3 is removed from the circuit to control the build-up of these minor impurities in the process.

In the electrowinning circuit about 57.5% of the copper is recovered per pass as a high purity product from the main electrowinning cells. This equates to about 84.5% of total copper production. About 27.5% of the copper entering the circuit is recycled through the anolyte compartment and the turboaerators to the leaching circuit. About 35% of the catholyte flow, containing the remaining 15% of the copper, is bled to a stripper cell where about 10.5% of the total copper is recovered as a relatively impure product. The remaining 4.5% of the copper is precipitated with H_2S generated in the preleach circuit, and the catholyte bleed stream containing the Cu_2S precipitate is recycled to provide the solution make up requirements for preleaching. The size of the catholyte bleed stream is governed primarily by the requirement to maintain an iron and water balance in the process.

The preleach step generates much more H_2S than can be used for copper precipitation from the catholyte bleed stream, even if the stripper cell was by-passed and all of the copper from this stream was precipitated. To minimize the problems and costs associated with disposal of this hazardous gas, the H_2S was oxidized to elemental sulphur in the secondary turboaerator by generating additional ferric ion. If desirable this elemental sulphur product, which is high purity and amounts to about 29.5% of the total sulphur produced, can be recovered separately by filtration prior to recycle of the STA solution to the oxidation leach.

The final leach residue from the oxidation leach consists of pyrite, elemental sulphur and gangue. Approximately 60 to 70% of the elemental sulphur is recovered by hot pressure filtration at 150°C. Sulphur recovery depends on the loss in the filter cake, which can range from 30-40% of the residue weight. The elemental sulphur product contains 1200

ppm Se, and is therefore not readily marketable. Essentially all of the gold remains with the pyrite, gangue and unreacted sulphides in the filter cake from sulphur filtration. The preferred process for gold recovery from this sulphide filter cake has not been finalized and is the subject of a further study.

The Woodall-Duckham process (13), which is in commercial use in the steel industry for processing acid pickle liquors, was selected for the oxyhydrolysis step in the GCM process. Although the process is very efficient for conversion of iron chloride to hematite and HCl, the efficiency of conversion of the nonferrous components (primarily Zn, Mg and Ca) to sulphates by adding sulphuric acid has not been investigated on a large scale. About 18% of the total chloride required for the process is tied up with these nonferrous components. High conversion efficiencies will therefore be necessary to minimize HCl make up requirements and costs, and to facilitate the future recovery of chloride free zinc sulphate from the hematite residue.

Further information is available in Appendix A - Process Design Criteria and Appendix B - Mass Balance.

4. ENGINEERING STUDY - DESCRIPTION & SCOPE

The following is an outline of the scope of work for the engineering study:

a) using the information provided by Bacon, Donaldson and Associates, prepare a more detailed engineering design for the GCM Process;

b) improve the accuracy of the capital and operating costs for the Process;

c) determine the benefits of implementing the GCM hydrometallurgical copper technology at Flin Flon;

d) contact acid pickle liquor plants in Canada and the USA to obtain technical information and engineering assistance to more closely define the capital and operating costs for oxyhydrolysis; and

e) prepare a final report on this phase of the study with conclusions and recommendations regarding the next phase of development (i.e. piloting, additional laboratory work, etc.)

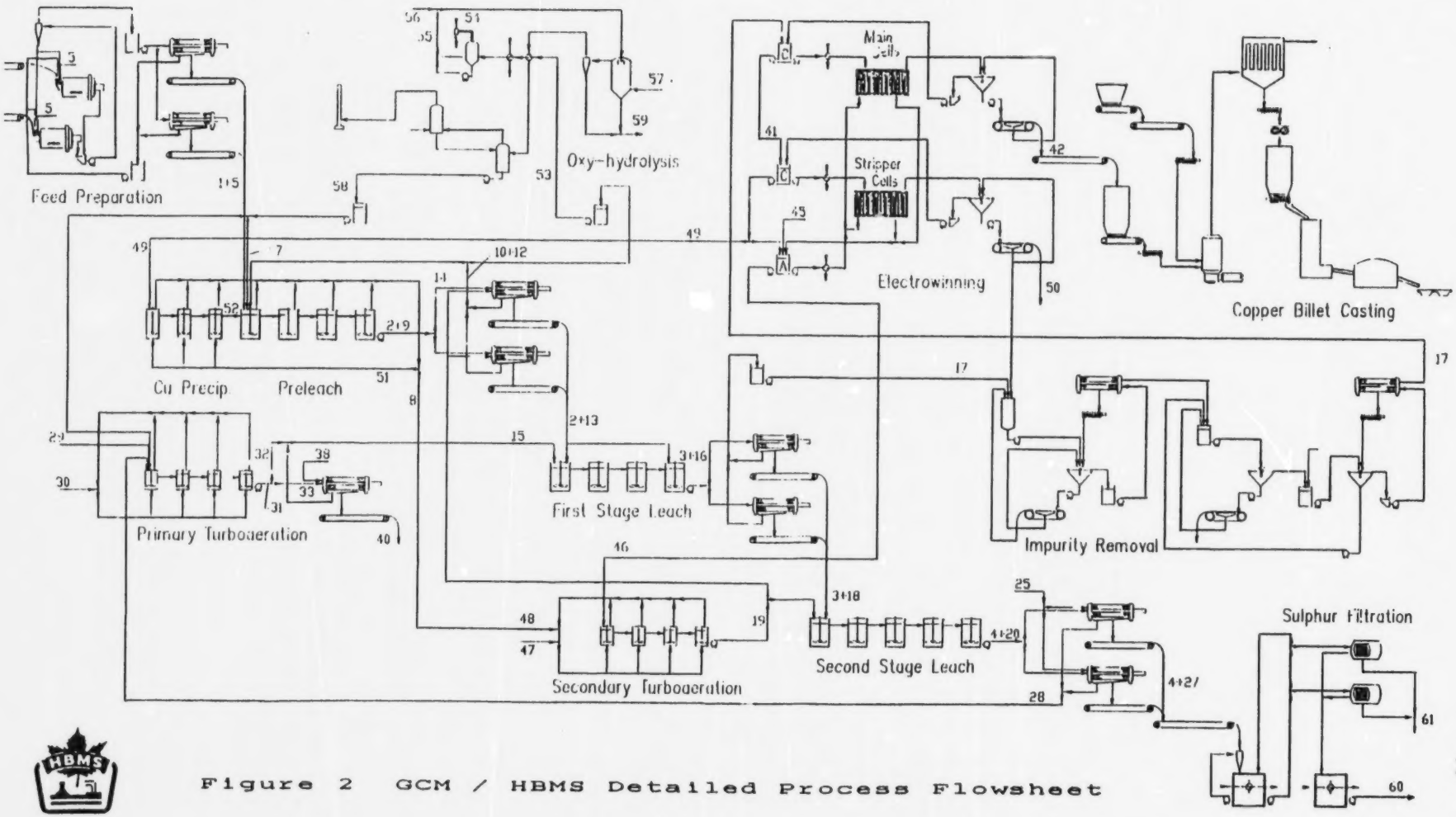


Figure 2 GCM / HBMS Detailed Process Flowsheet

The preliminary flowsheet provided by BDA (Figure 4, Appendix B) was expanded to include concentrate unloading and blending as well as facilities required to produce final products - i.e. cast copper billets. To minimize capital costs and heat losses in the leach circuit, thickeners were removed from the final process flowsheet.

The resulting detailed flowsheet (Figure 2) was developed in consultation with Mr. A. Fletcher, consultant, and Mr. W.J.S. Craigen (CANMET). Equipment design and materials of construction were also determined in consultation with Mr. A. Fletcher.

Based on the BDA design criteria and material balances (Appendices A and B respectively), equipment was selected and sized, and prices obtained from suppliers if recent cost data were not available at HBM&S.

Some minor revisions to the original BDA material balances were necessary. These were primarily related to sulfuric acid requirements and Jarosite formation in the primary turboaerator (PTA) circuit, and to the ferric/cupric ratios in the leach and turboaeration circuits, where it was assumed that cupric is the most stable oxidized form (based on thermodynamic considerations). These revisions are reflected in the mass balances shown in Appendix B, Table VI.

A trip to Hamilton (Stelco) and Pittsburgh (Perox) by CANMET and Hudbay representatives was undertaken to gain operating, engineering and cost information on Woodall-Duckham acid pickle liquor recovery plants, and to determine the applicability of this process for oxyhydrolysis of the GCM process bleed stream. As a result of this visit, some further changes were made in the BDA mass and energy balances for the oxyhydrolysis circuit. Capital and operating costs for oxyhydrolysis were based primarily on data from Perox (14).

To more accurately assess the thermal energy requirements for the GCM process, detailed energy balances were calculated for the preleach, copper precipitation, reduction leach, oxidation leach, primary turboaerator, secondary turboaerator and evaporator circuits. These calculations included heats of reaction based on the process chemical reactions (Appendix A, Table II), estimated heat losses in the leaching and turboaeration circuits and the utilization of waste heat in the oxyhydrolyzer off-gases for water removal in the evaporator. The effect of Jarosite formation in the PTA and redissolution in the oxidation leach was also determined. Oxyhydrolyzer energy requirements were based

primarily on data provided by Perox (14). For the purification/electrowinning circuits it was assumed that heat generated from electrical inefficiency in electrowinning would more than compensate for heat losses in the circuits.

Total energy requirements for the GCM process, as applied to HBM&S copper concentrates, are shown in Appendix C, Table VII, with thermal energy requirements broken down into leaching, turboaeration, evaporation and oxyhydrolysis circuits. In calculating steam requirements for water evaporation, a single effect evaporator is assumed, but full credit is applied for the excess heat generated in the two turboaerators.

5. PLANT LAYOUT AND LOCATION

The process was divided into the following sub-groups which are depicted on the detailed process flowsheet in Figure 2.

Feed Preparation
Pre-Leach
First and Second Stage Leaching
Impurity Removal
Electrowinning
Copper Precipitation
Primary and Secondary Turboaeration
Oxyhydrolysis
Sulphur Filtration
Copper Billet Casting

A "greenfields" site in Flin Flon was assumed - with no exact location specified. The site would be relatively flat with no major earth work or rock excavation being required.

A simplified layout of the plant is provided in Figure 3. The area of land required for this installation is approximately 7.5 hectares.

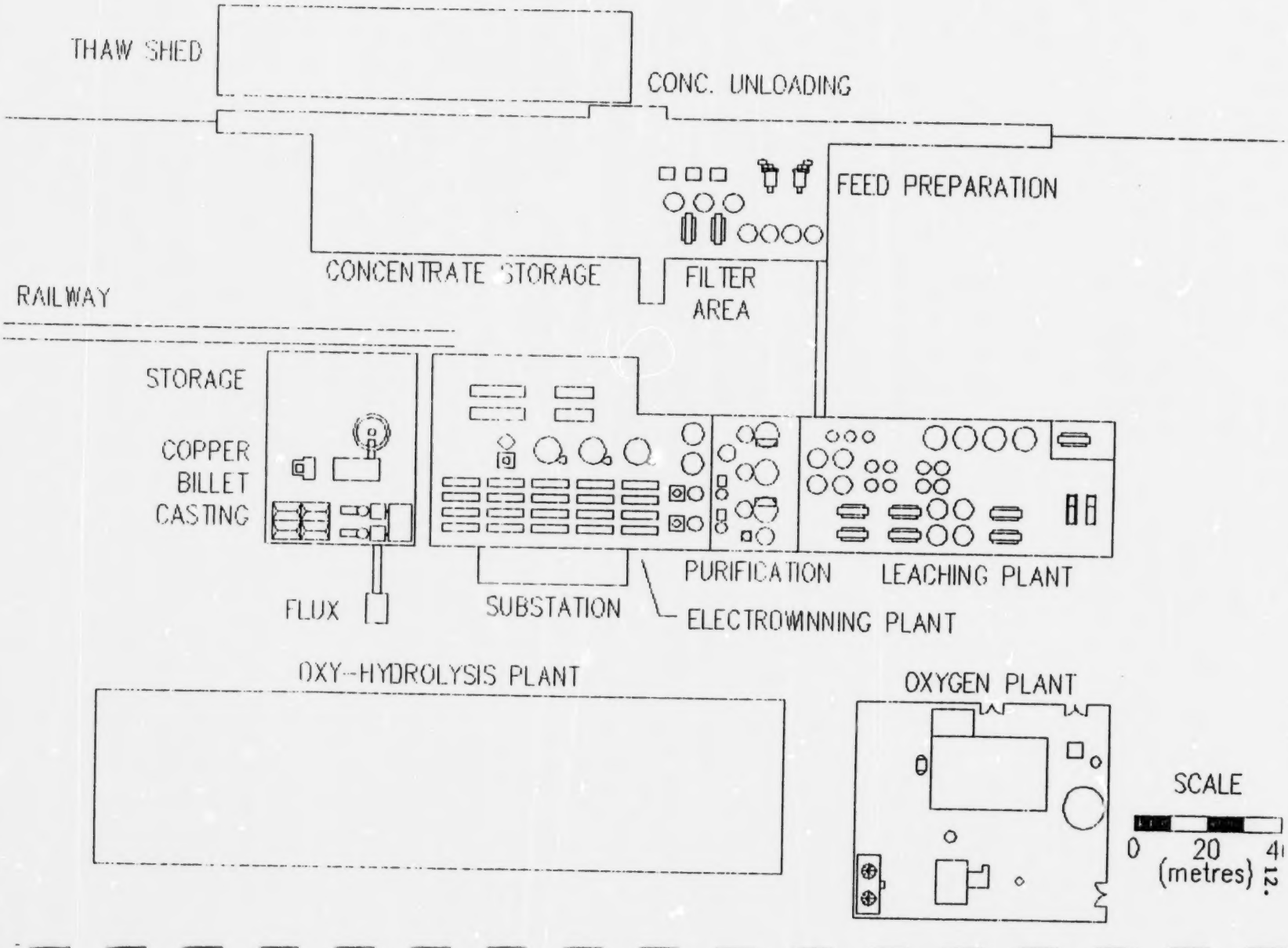


Figure 3 Plant Layout

6. CAPITAL COSTS

The capital cost to construct a plant capable of producing 75000 tonnes per year of copper is $157 071 000. This cost is expected to have an accuracy of +20% - 10%. Capital costs are expressed in 1986 dollars and are summarized on Table VIII, Appendix D. Capital costs are broken down into more detail on Table IX in the same appendix.

The preliminary flowsheet prepared by BDA was modified to include concentrate handling prior to feed preparation and copper billet casting after electrowinning. Capital costs reflect this change in scope. A comparison of the BDA and HBM&S capital cost estimates is shown in Table X, Appendix D. The major differences are due to the aforementioned changes in scope, as well as updated costs for sulphur filtration and oxyhydrolysis. The contingency was adjusted to reflect the amount of feasibility engineering completed in the study.

In the breakdown of capital costs, materials and spares have been estimated at 32.0% of the total. The equipment requirements and pump list are described on Tables XI and XII respectively in Appendix D. Engineering and project management are 10% of the direct plant costs and the contingency is 20%.

7. OPERATING COSTS

A breakdown of HBM&S and BDA operating cost estimates for the GCM process are shown for comparison in Table XIII, Appendix E. The HBM&S estimate, at 17.76¢/lb copper, is slightly below the BDA estimate of 20.21¢/lb. BDA costs are based on a throughput of 1000 tonnes/day of concentrate at 20.5% Cu, which is typical of HBM&S Copper concentrate with analyses as shown in Table I. HBM&S costs are based on 848 tonnes/day of concentrate at 23.7% Cu, which is typical of the present blend of HBM&S and custom concentrates fed to the copper smelter.

The change in concentrate grade and throughput has minimal effect on GCM process operating costs, because cuprous ion solubility governs solution flows and copper output is about the same for both feeds. Major factors which have affected the HBM&S cost estimates are:

- Slightly decreased reagent costs, after the mass balance was revised and sulfuric acid requirements were decreased, and reagent consumption was adjusted to reflect the higher concentrate grade.

- Increased utility costs due to higher fuel unit costs and a more accurate assessment of the thermal energy requirements of the process.

- Lower direct labour and overhead costs, as these were made to conform to HBM&S plant operating practice, and overhead was adjusted to reflect the proportion carried by the present copper plant.

Operating costs are based on unit costs and requirements shown in Table XIV, Appendix E. Unit costs, with the exception of fuel oil and labour, are essentially the same as those used by BDA, and reflect costs at Flin Flon.

8. FINANCIAL EVALUATION

The financial evaluation determines the impact of the GCM process plant operating in conjunction with the Zinc Pressure Leach (ZPL) facility. A summary of the evaluation appears on Table XV, Appendix F.

The base case for this evaluation is Hudbay's 1986 operating budget forecast modified to include Zinc Pressure Leach (15). The mode of operation of the copper smelter remains unchanged, however the slag fuming furnace is shut down after the ZPL process is introduced. The comparison of the base case with the GCM option assumes that concentrate inputs are the same, and that product outputs are dependent upon recoveries. For the base case, overall zinc recovery is estimated at 86.9%, based on 97% recovery in the ZPL process and zero recovery from the copper concentrate. For the GCM option, overall zinc recovery is estimated at 95.4% based on 97% recovery in the ZPL and 83.8% recovery from the Cu concentrate. The latter figure is based on leaching and precipitating zinc from the GCM process hematite residue.

Although the BDA material balances show a copper recovery of 99%, for the financial evaluation a recovery of 98% has been assumed, or an increase of 2.5% over the base case. It is also assumed that copper, silver and gold recovery from the zinc concentrates is the same for both base case and GCM options, since ZPL plant copper and sulfide residues will be recycled to either the copper smelter (base case) or the GCM process, for further treatment and recovery.

For the GCM option, overall silver recovery is slightly less than the base case due to silver losses in the copper product and unrecoverable silver in the leach residue. Gold recovery however, is slightly higher, since it is assumed that 100% of the gold ends up in the GCM and ZPL process residues, and could be recovered at a custom processor.

Metals distribution in the GCM process is shown in Table XVI, Appendix F.

With the implementation of the GCM process, changes in product output, revenues and operating costs are calculation from the base case. Incremental changes in outputs and revenues for the GCM option versus the base case are shown in Table XVII, Appendix F. Revenues are based on first quarter 1987 metal prices, shown in Table XVIII, Appendix F. Precious metal revenues from the GCM and ZPL process leach residues are based on the processing,refining and shipping charges also shown in Table XVIII.

For the GCM process, residue weight has been reduced from the figure reported in the BDA material balances to reflect the higher grade of concentrate (23.7% Cu versus 20.5% Cu). Residue weight was calculated based on the assumption that the major diluent components (pyrite, pyrrhotite and gangue) are all reduced in propertion to their original concentrations (shown in Table I for the lower grade concentrate). Similarly sulfur output and revenue has been adjusted to reflect a lower loss in the leach residue (sulfur loss is proportional to leach residue weight), and the amount required to supply H_2SO_4 to the GCM process (8200 mtpy).

Incremental changes in operating costs for the GCM option versus the base case are shown in Table XIX, Appendix F. Although operating costs for the GCM option are higher, these are more than offset by the decrease in copper and PM refining charges, resulting in a net operating cost savings of about $5.6 million/y.

The financial analysis shown in Table XV indicates a total potential benefit for the GCM option of about $7.3 million/y. About 23% of this benefit is due to increase revenues resulting from sulfur sales and higher recoveries of copper and zinc. The remaining 77% is due to operating cost savings resulting from lower copper and PM refining charges.

9. CONCLUSIONS

Based on the total net benefit and pretax cash flows, the financial evaluation indicates a capital payback time for the GCM option of about 21 years, or an ROI of only about 4.7% even if interest on borrowed capital is neglected. This is not an attractive investment for HBM&S, particularly since operating costs, on a copper throughput basis, are higher for the GCM option, and net benefits are primarily dependent upon unproven technical factors related to the production of high purity copper and saleable sulfur and zinc by-products.

The major factors affecting economic viability are:

a) Copper product purity:

The copper product must be equivalent to cathode quality to ensure the substantial savings shown for copper refining costs.

b) Sulfur marketability:

The sulfur produced in both the GCM and ZPL processes contains high levels of certain impurities (such as Se) which could affect the price and marketability of the sulfur product.

c) Zinc recovery and efficiency of HCl regeneration in oxyhydrolysis:

The credit for zinc recovery and the HCl component of total GCM process operating costs are very much dependent upon the efficiency of HCl recovery from the nonferrous components in the oxyhydrolysis step. If HCl cannot be efficiently recovered from these components, GCM process operating costs could increase by over 6¢/lb copper and chloride-free zinc, for recovery in the ZPL process, could not be produced.

d Gold revenue:

The treatment charges and shipping costs associated with gold recovery from the GCM and ZPL process residues, results in a 30% decrease in net gold revenue to HBM&S. Methods for improving elemental sulfur separation from the leach residues could significantly reduce residue weight (up to 30%) and thereby reduce treatment and shipping costs and increase gold revenues to HBM&S. An on-site treatment facility for gold recovery could also improve gold revenues. These options are particularly important for the economic processing of lower grade copper concentrates, such as the HBM&S concentrate on which

the BDA material balances were based, where residue weights are substantailly higher.

10. RECOMMENDATIONS

Of the four factors listed above which have a major effect on the financial evaluation of the GCM option, the largest is the loss of gold revenue ($14 million/y). If this loss in revenue could be reduced through the development of an on-site gold extraction and recovery system, the benefits associated with the GCM option could increase up to about 13% ROI, or a 7½ year capital write-off (neglecting interest charges). It is desirable therefore to investigate the most economical methods for recovering gold from the GCM and ZPL residues to maximize the return to HBM&S. This investigation has already been initiated in a separate project under the Manitoba MDA.

Even if the gold revenue problem is successfully resolved, the GCM process cannot be commercially implemented until it has been shown that the other major criteria affecting economic viability can be successfully met. To accomplish this, continuous pilot scale testwork on the GCM process is required to determine:

a) The purity and subsequent value of the copper product when operating the process under optimum conditions for producing high purity copper.

b) The efficiency of HCl regeneration from non-ferrous impurities (Zn, Ca, Mg) by H_2SO_4 addition during oxyhydrolysis, and the feasibility of producing a chloride-free zinc product. Alternately, H_2SO_4 could be considered in place of HCl for the preleach which would facilitate the recovery of a chloride-free zinc product.

c) Better methods to separate sulphur from the leach residue, to maximize sulphur recovery and minimize volume of residue which has to be treated for gold recovery.

d) A method to improve the purity and saleability of the sulphur by-product.

e) The engineering aspects of the process, particularly the optimum design of the electrolytic cells to facilitate the recovery of the particulate copper product.

Items (c) and (d) will benefit both the GCM and ZPL processes. The cost of piloting the GCM process will probably be at least $5 million. Nevertheless, it is recommended that pilot scale testing directed towards verifying and improving the economic viability of the process be considered, based on the potential to economically eliminate SO_2 emissions, but subject to a favourable resulution of the gold revenue loss problem.

11. REFERENCES

1. EMR Canada; Canada's Nonferrous Metals Industry: Nickel and Copper A Special Report, 1984.

2. EMR Canada; Canada's Nonferrous Metals Industry: Lead and Zinc A Special Report, 1985.

3. Craigen W.J.S. and King D.E.C.; Inco Sudbury Technological Profile. CANMET/OTT 84-04 (TR); June 1984.

4. Craigen W.J.S., King D.E.C. and Wong W.S.H.: Falconbridge Technological Profile CANMET/OTT 84-05 (TR); January 1985.

5. Craigen W.J.S. and King D.E.C.; Inco Thompson Technological Profile; CANMET MRP/MSL 85-87 (TR); June 1985.

6. Craigen W.J.S. and King D.E.C.; HBM&S Technological Profile; CANMET MRP/MSL 85-48 (TR); June 1985.

7. Chalkely M.E.; Masters I.M. and Weir D.R.; Miniplant Pressure Leaching of Zinc. CIM Conference of Metallurgists, 13th Annual Hydrometallurgy Meeting; August 1983; Edmonton, Alta.

8. Beattie M.J.V., Craigen W.J.S. and Sarkar K.M.; The Great Central Mines Hydrometallurgical Copper Process; CANMET MRP/OTT 84-08 (j); October 1984.

9. Craigen W.J.S. and Beattie M.J.V.; Economics of Chemical Methods for Copper Extraction and Refining; CANMET MRP/MSL 85-85 (TR); June 1985.

10. Craigen W.J.S.; GCM Hydrometallurgical Copper Process: Evaluation and Recommendations for Commercial Application; CANMET MRP 80-02 (CF); June 1980.

11. Craigen W.J.S. and Kelly F.J.; GCM Hydrometallurgical Copper Process: Evaluation of Modified Process; CANMET MRP/OTT 82-01 (CF), November 1982.

12. Bacon Donaldson and Associates; Bench Scale Evaluation of the Application of the GCM Hydrometallurgical Copper Process to HBM&S Copper Concentrates; Volumes I, II and III; June 1986.

13. Conners A; Hydrochloric Acid Regeneration as Applied to the Steel and Mineral Processing Industries; CIM Bulletin; February 1985, p. 75.

14. Wadhawan S; Perox Inc. Pittsburgh, Penn.; Private Communication.

15. HBM&S: Zinc Pressure Leach Technical and Capital Cost Report; November 1985.

APPENDIX A

PROCESS DESIGN CRITERIA

TABLE I Concentrate Analysis and Mineralogy

Concentrate Analysis and Mineralogy

Concentrate Analysis, %			
Cu	20.49	Co	0.052
Fe	32.08	Ni	0.003
S	34.49	Cd	0.008
Zn	2.88	Mo	0.003
Pb	0.17	MnO	0.03
As	0.13	Se	0.026
Sb	0.047	Te	0.0038
Al_2O_3	0.68	Bi	<0.01
MgO	0.62		
CaO	0.67	F	0.014
SiO_2	3.84		
Ag	87.5 g/tonne	Au	7.6 g/tonne

Concentrate Mineralogy

$CuFeS_2$	59.2
FeS	10.3
FeS_2	16.1
ZnS	4.3
Minor + Gangue	10.0

TABLE II

GCM PROCESS CHEMICAL REACTIONS

PRELEACH

$$ZnS + 2\ HCl \longrightarrow ZnCl_2 + H_2S$$
$$PbS + 2\ HCl \longrightarrow PbCl_2 + H_2S$$
$$CaCO_3/MgCO_3 + 4\ HCl \longrightarrow CaCl_2/MgCl_2 + 2\ H_2O + 2\ CO_2$$

REDUCTION LEACH

$$CuFeS_2 + 3\ FeCl_3 \longrightarrow CuCl + 4\ FeCl_2 + 2\ S^\circ$$
$$CuFeS_2 + 3\ CuCl_2 \longrightarrow 4\ CuCl + FeCl_2 + 2\ S^\circ$$

OXIDATION LEACH

$$CuFeS_2 + 4\ FeCl_3 \longrightarrow CuCl_2 + 5\ FeCl_2 + 2\ S^\circ$$
$$Fe_2O_3 + 6\ HCl \longrightarrow 2\ FeCl_3 + 3\ H_2O$$

ELECTROWINNING

CATHOLYTE: $H^+ + CuCl \longrightarrow Cu^\circ + HCl\ (L)$

ANOLYTE:
$$FeCl_2 + HCl \longrightarrow FeCl_3 + H^+$$
$$CuCl + HCl \longrightarrow CuCl_2 + H^+$$

COPPER PRECIPITATION

$$2\ CuCl + H_2S \longrightarrow Cu_2S + 2\ HCl\ (L)$$

PRIMARY TURBOAERATION

$$2\ FeCl_2 + \tfrac{1}{2}\ O_2 + 2\ H_2O \longrightarrow Fe_2O_3 + 4\ HCl\ (L)$$
$$2\ FeCl_3 + 3\ H_2O \longrightarrow Fe_2O_3 + 6\ HCl\ (L)$$
$$2\ FeCl_2 + \tfrac{1}{2}\ O_2 + 2\ HCl \longrightarrow 2\ FeCl_3 + H_2O\ (L)$$

SECONDARY TURBOAERATION

$$2\ FeCl_2 + \tfrac{1}{2}\ O_2 + 2\ HCl \longrightarrow 2\ FeCl_3 + H_2O\ (L)$$
$$2\ FeCl_3 + H_2S \longrightarrow 2\ FeCl_2 + 2\ HCl\ (L) + S^\circ$$

EVAPORATION

$$2\ FeCl_2 + \tfrac{1}{2}\ O_2 + 2\ HCL \longrightarrow 2\ FeCl_3 + H_2O\ (G)$$

OXYHYDROLYSIS

$$2FeCl_2 + \tfrac{1}{2}\ O_2 + 2\ H_2O \longrightarrow Fe_2O_3 + 4\ HCl\ (G)$$
$$2\ FeCl_3 + 3\ H_2O \longrightarrow Fe_2O_3 + 6\ HCL\ (G)$$] 82%

$$ZnCl_2 + H_2SO_4 \longrightarrow ZnSO_4 + 2\ HCl\ (G)$$ 11%

$$CaCl_2/MgCl_2 + 2\ H_2SO_4 \longrightarrow CaSO_4/MgSO_4 + 4\ HCl\ (G)$$
$$PbCl_2 + H_2SO_4 \longrightarrow PbSO_4 + 2\ HCl\ (G)$$] 7%

TABLE III

EQUIPMENT DESIGN CRITERIA

The following is the recommended process design criteria for the unit operations of the GCM process to treat HBM&S copper concentrates:

4.1 CONCENTRATE PREPARATION

Concentrate Regrind: 80% passing 28 microns

4.2 PRELEACH AND COPPER PRECIPITATION

4.2.1 Copper Precipitation

Temperature: 90°C
Retention Time: 2 hours
HCl Concentration: 0.75 - 1.0 molar
Fe Concentration: 139 gpl
% Solids: initial 0%
final 1.3%

4.2.2 Preleach

Temperature: 90°C
Retention Time: 4 hours
HCl Concentration: initial 1.0 molar
final 0.5 molar
Fe Concentration: initial 149 gpl
final 156 gpl
% Extraction: Pyrrhotite 95%
Zn 92%
Pb 85%
% Solids: initial 23%
final 20%
Filtration Rate (based on vacuum tests):
(Note: No Cu_2S solids present)

9.8 kg solids/m^2 h (14% solids feed)
161 kg solids/m^2 h (65% solids feed)
77 max % solids cake

TABLE III - Cont'd

4.3 **LEACHING (1ST AND 2ND STAGE)**

Temperature: 105° - $108^{\circ}C$
Retention Time: 1st Stage - 2 hours
2nd Stage - 4 hours
Overall % Extraction: Cu > 99%
Zn > 98%
Pb > 98%
Ag 93 to 98%
Pyrrhotite > 99%
Solution Compositions: gpl

	1st Stage		2nd Stage	
	initial	final	initial	final
Fe_T	156	138	115.9	161.3
Fe^{3+}		1	84.8	26.4
Cu_T		110	50.6	87.5
Cu^{+}		110	20.2	87.5
HCl	18		26.5	

		1st Stage	2nd Stage
% Solids:	initial	15.9%	15.8%
	final	13.3%	11.3%

2nd Stage Filtration Rate (based on vacuum tests):
256 - 420 kg solids/m^2 h (35.3% solids feed)

Filter Cake Wash: 58% max % solids cake
2 displacements

TABLE III – Cont'd

4.4 TURBOAERATION

4.4.1 Primary (PTA)

Temperature: 90° - 100°C
Retention Time: 8 hours
Solution Composition: gpl

	initial	final
Cu	80	80
Fe_T	147	110
Fe^{3+}	0	100
SO_4	3 gpl	

O_2 Flow: 0.25 m^3/min m^3 solution (O_2 at 25 C)
O_2 Consumption: 0.27 kg/min m^3 solution
Equivelent Ferric Production: 1.9 kg Fe^{3+}/min m^3
% of Total Fe in PTA Feed Precipitated: 25%
% of Fe Precipitated Recycled to 1st Stage: 33%
% Solids Feed to PTA Filter: 4.3%
Filtration Rate (based on vacuum tests):
1.4 - 12.2 kg solids/m^2 h
60% max % solids cake
Agitation: 0.30 kW/m^3 (1.5 hp/1000 US gal)

4.4.2 Secondary (STA)

Temperature: 90° - 100°C
Retention Time: 4 hours
Solution Composition: gpl

	initial	final
Cu	38	37
Fe_T	113	111
Fe^{3+}	50	104

O_2 Flow: 0.25 m^3/min m^3 solution (O_2 at 25 C)
O_2 Consumption: 0.22 kg/min m^3 solution
H_2S Oxidation: 48 tonnes/d
Elemental S Produced: 45.5 tonnes/d
Agitation: 0.30 kW/m^3 (1.5 hp/1000 US gal)

TABLE III – Cont'd

4.5 **PURIFICATION**

4.5.1 **Gold Removal**

Copper Powder: 10 gpl

4.5.2 **Silver Removal**

Temperature: 90°C
Retention Time: 30 minues per stage (2 stages)
Cuprous Iodide Concentration: 5 gpl

4.6 **ELECTROWINNING**

4.6.1 **Main Cells**

Temperature: 80°C
Solution Composition: gpl

	Catholyte		Anolyte	
	initial	final	initial	final
Cu	110	47	47	38
Fe_T	139	139	139	113
Fe^{3+}	1			51
HCl		37	124	69

Electrodes: Cathode: Titanium
Anode: Ru-coated DSA
Electrode Size: 1.5 m^2
Electrode Spacing: 2.5 cm_2 (membrane centered)
Current Density: 1000 A/M^2
Current Efficiency: $>$ 99%
Power Consumption: 0.57 kWh/kg Cu
Copper Deposition Rate: 2.3 – 2.4 kg/m^2 h
% Copper Recovered per Pass: 57.5%
% Copper Recycled to Leach: 27.5% as $Cu\ Cl_2$
4.5% as Cu_2S

TABLE III - Cont'd

4.6.2 **Stripper Cells**

% Copper recovered from stripper: 10.5%
Temperature: 80°C
Solution Composition: gpl

	Catholyte	
	initial	final
Cu	47	14
Fe_T	139	139
Fe^{3+}		
HCl	37	56

Current Density: 400 A/m^2
Current Efficiency: 99%
Power Consumption: 0.53-0.54 kWh/kg Cu
Copper Deposition Rate: 0.94 kg/m^2 h

4.7 **EVAPORATION/OXYHYDROLYSIS**

Stream Composition:

	Evaporator		Oxyhydrolysis	
	initial	final	Feed	Product Gas
	gpl		%	
$FeCl_2$	354	522	30.0	
$FeCl_3$	0	147	8.5	
HCl	18	0	0	18.6
Zn	26	47	2.7	
Cl_T	240	463	26.7	
H_2O				34.4

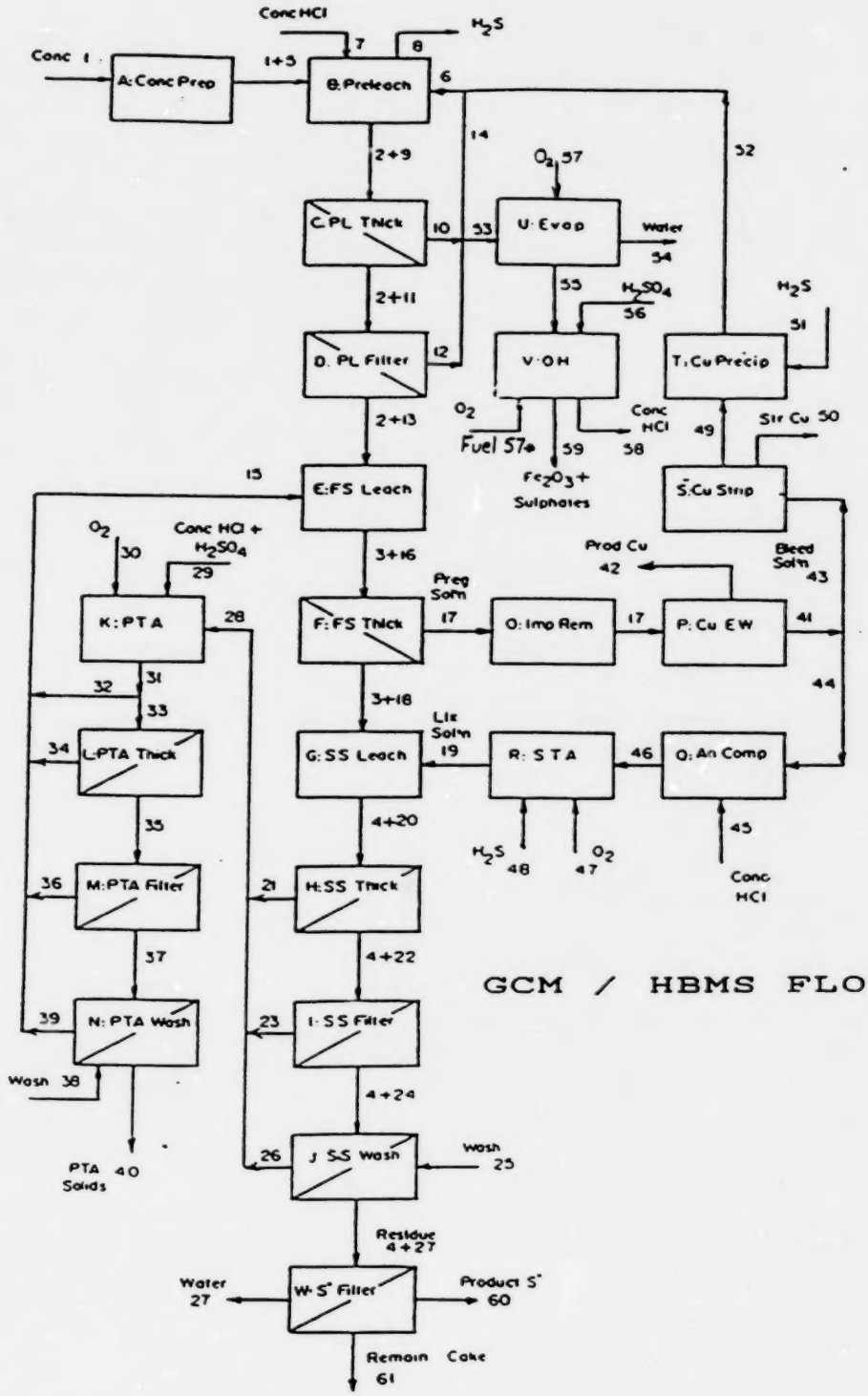
Fe Removal Required (for Fe balance): 7134 kg/h
H_2O Removal Required (for H_2O balance): 20 245 kg/h
H_2SO_4 Required: Stoichiometric for Zn, Ca, Mg and Pb

4.8 **ZINC RECOVERY**

% Zn in hematite: 8.34%

APPENDIX B

MASS BALANCE



GCM / HBMS FLOWSHEET

FIGURE 4 Flowsheet Stream Identification

TABLE IV List of Unit Operations

A: Concentrate Preparation
B: Acid Preleach
C: Preleach Thickener
D: Preleach Filter
E: First Stage Leach
F: First Stage Thickener
G: Second Stage Leach
H: Second Stage Thickener
I: Second Stage Filter
J: Second Stage Wash
K: Primary Turboaerator
L: Primary Turboaerator Thickener
M: Primary Turboaerator Filter
N: Primary Turboaerator Wash
O: Impurity Removal
P: Copper Electrowinnning
Q: Anode Compartments
R: Secondary Turboaerator
S: Copper Stripper
T: Copper Precipitation
U: Evaporator
V: Oxyhydrolysis
W: Sulphur Filter

TABLE V List of Process Flows

1. Concentrate (S)
2. Preleached Concentrate (S)
3. First Stage Leach Residue (S)
4. Second Stage Leach Residue (S)
5. Water with Concentrate (L)
6. Preleach Feed Solution (L)
7. Conc HCl to Preleach (L)
8. Hydrogen Sulphide Produced (G)
9. Preleach Product Solution (L)
10. Preleach Thickener Overflow (L)
11. Solution in Preleach Thickener Underflow (L)
12. Solution Taken off on the Preleach Filter (L)
13. Solution in the Preleach Filter Cake (L)
14. Preleach Recycle (L)
15. First Stage Leach Feed Solution (L)
16. First Stage Product Solution (L)
17. Pregnant Solution to Electrowinning (L)
18. Solution in the First Stage Thickener Underflow (L)
19. Lixiviant Solution to the Second Stage Leach (L)
20. Second Stage Product Solution (L)
21. Second Stage Leach Thickener Overflow (L)
22. Solution in the Second Stage Thickener Underflow (L)
23. Solution Taken off on the Second Stage Leach Filter (L)
24. Solution in the Second Stage Leach Filter Cake (L)
25. Second Stage Wash Solution (L)
26. Solution from the Second Stage Wash (L)
27. Solution in the Second Stage Washed Cake
28. Primary Turboaerator Feed (L)
29. Hydrochloric and Sulphuric acid to the Primary Turboaerator
30. Oxygen in the Primary Turboaerator (G)
31. Primary Turboaerator Product Slurry (SL)
32. Primary Turboaerator Slurry to the First Stage Leach (SL)
33. Primary Turboaerator Slurry to the Thickner (SL)
34. Primary Turboaerator Thickener Overflow (L)
35. Primary Turboaerator Thickener Underflow (SL)
36. Primary Turboaerator Filtrate (L)
37. Primary Turboaerator Filter Cake (SL)
38. Primary Turboaerator Wash Solution (L)
39. Solution from the Primary Turboaerator Wash (L)
40. Washed Primary Turboaerator Filter Cake (SL)
41. Solution from Electrowinning (L)
42. Product Copper (S)
43. Solution Bleed (L)
44. Anolyte Feed (L)
45. Concentrated HCl to the Anode Compartments (L)
46. Secondary Turboaerator Feed (L)
47. Oxygen to the Secondary Turboaerator (G)
48. H_2S to the Secondary Turboaerator (G)
49. Solution to Copper Precipitation (L)
50. Stripped Copper (S)
51. H_2S to Copper Precipitation (G)
52. Stripped Bleed Solution (L)
53. Evaporator Feed (L)
54. Evaporated Water (L)
55. Oxyhydrolysis Feed (L)
56. Sulphuric Acid to Oxyhydrolysis (L)
57. Air + Fuel to Oxyhydrolysis and O2 to evaporator (G)
58. Product Concentrated HCl (L)
59. Hematite and Metal Sulphates from Oxyhydrolysis (S)
60. Product Sulphur from Sulphur Filtration(S)
61. Remaining Residue Cake from Sulphur Filtration(S)

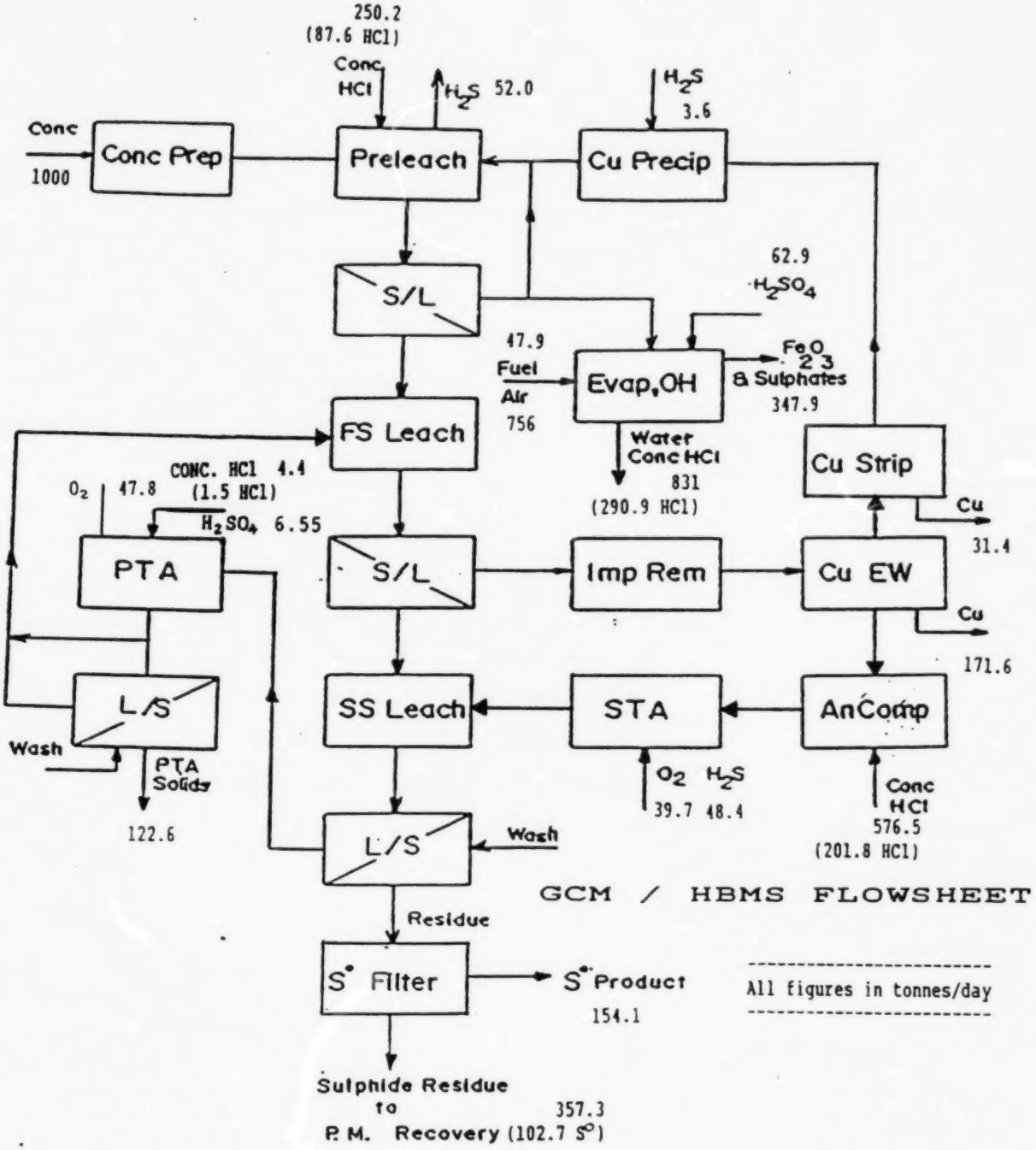


FIGURE 5 Major Stream Flows

TABLE VI

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows — Solid Flows

Stream	Conc	PL Prod	PS Prod	Residue	Prod S	Rem Cake	Prod Cu	Str Cu	PTA Sol	OH Sol
Phase	S	S	S	S	S	S	S	S	S	S
No.	1	2	3	4	60	61	42	50	40	59
t/day	1000.0	866.4	710.8	511.4	154.1	357.3	171.6	31.4	112.0	345.7
kg/min	694.4	601.7	493.6	355.2	107.0	248.2	119.1	21.8	77.8	240.1
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
CuFeS2	411.1	411.1	205.6	4.1		4.1				
ZnS	29.9	2.4								
PbS	1.-	0.2								
FeS	71.6	3.6								-70.1
FeS2	112.2	112.2	112.2	112.2		112.2				Sulphates
Cu2S		11.7								
S			76.3	178.3	107.0	71.3				
MgO++	7.60									
Cu							119.1	21.8		
Gangue	60.6	60.6	60.6	60.6		60.6				
Fe203			33.3						66.6	170.0
H4Fe6(SO4)4(OH)14			5.58						11.2	
Cu	142.4	151.7	71.2	1.4		1.4	119.1	21.8		
Zn	20.0	1.6								20.0
Pb	1.2									1.2
Fe	222.8	179.6	138.0	53.5		53.5			50.5	118.9
Mg+	4.6									4.6
S	239.7	208.1	208.1	239.7	107.0	71.3				
O	2.2		10.0							
Gangue	60.6	60.6	60.6	60.6		60.6				
Cl	6.0			1.0		1.0			1.0	3.0
	%	%	%	%	%	%	%	%	%	%
CuFeS2	59.20	68.33	41.35	1.16		1.66				
ZnS	4.30	0.40								
PbS	0.20									
FeS	10.31	0.59								
FeS2	16.15	18.64	22.56	31.58		45.19				
Cu2S		1.94								-29.69
S			15.35	50.21	100.00	28.75				Sulphates
MgO++	1.10									
Cu							100.00	100.00		
Gangue	8.72	10.06	12.18	17.05		24.40				
Fe203			6.70						78.18	70.80
H4Fe6(SO4)4(OH)14			1.12						21.82	
Cu	20.50	25.21	14.32	0.40		0.57	100.00	100.00		
Zn	2.89	0.27								8.34
Pb	0.17									0.50
Fe	32.08	29.85	27.77	15.05		21.54			59.24	49.52
Mg+	0.66									2.32
S	34.52	34.58	41.85	67.50	100.00	53.48				
O	0.32		2.01							
Gangue	8.72	10.06	12.18	17.05		24.40				
Cl				0.28		0.40			1.28	1.25

computer file: CCL-MASS.WK1 10-Jun-87 Mass balance by: BDA Revision by: HBMS

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows PTA Flows (1)

Stream	PTA Feed	PTA HCl	PT H2SO4	PTA O2	PTA Prod	TH BYPSS	TH Feed	PTATH OF	PTATH UF	PTAFL OF	PTAFL UF
Phase	L	L	L	G	SL	SL	SL	L	SL	L	SL
No.	28	29A	29B	30	31	32	33	34	35	36	37
t/day	4274.2	4.4	6.6	47.8	4310.4	1436.8	2873.5	2393.7	479.8	245.3	234.6
kg/min	2954.5	3.1	4.55	33.2	2993.3	997.8	1995.5	1662.3	333.2	170.3	162.9
l/min	2048.0	2.6			2051.8	683.9	1367.9	1185.7	182.2	121.5	60.7
sg	1.439	1.180			1.459	1.459	1.459	1.402	1.829	1.402	2.684
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
Cu++	56.32				164.2	54.7	109.5	94.9	14.6	9.7	4.9
Cu+	107.88										0.0
Zn++	10.40				10.4	3.5	6.9	6.0	0.9	0.6	0.3
Pb++	0.80				0.8	0.3	0.5	0.5	0.1	0.0	0.0
Fe++	302.80				165.9	55.3	110.6	96.0	14.6	10.0	4.6
Fe+++	0.00				61.2	20.4	40.8	35.3	5.5	3.6	1.8
Mg+	1.80				1.8	0.6	1.2	1.1	0.2	0.1	0.1
Cl-	522.20				525.1	175.0	350.1	303.4	46.6	31.1	15.5
SO4-											
HCl		1.10			-2.00	-0.70	-1.30	-1.10	-0.20	-0.10	-0.10
H2O	1945.60	2.0			1949.2	649.7	1299.5	1126.4	173.1	115.4	57.7
H2S											
Cu2S											
Fe2O3					99.9	33.3	66.6		66.6		66.6
H4Fe6(SO4)4(OH)14					16.74	5.58	11.16		11.16		11.16
H2SO4	2.20		4.55								
O2				33.2							
S											
	g/l	%	%	%	g/l	g/l	g/l	g/l	g/l	g/l	g/l
Cu++	27.50				80.02	80.02	80.02	80.02	80.02	80.02	80.02
Cu+	52.68										
Zn++	5.06				5.05	5.05	5.05	5.05	5.05	5.05	5.05
Pb++	0.40				0.40	0.40	0.40	0.40	0.40	0.40	0.40
Fe++	147.85				80.86	80.86	80.86	80.86	80.86	80.86	80.86
Fe+++	0.00				29.83	29.83	29.83	29.83	29.83	29.83	29.83
Mg+	0.90				0.89	0.89	0.89	0.89	0.89	0.89	0.89
Cl-	254.97				255.93	255.93	255.93	255.93	255.93	255.93	255.93
SO4-											
HCl		35.00			-0.95	-0.95	-0.95	-0.95	-0.95	-0.95	-0.95
H2O	950.00	65.00			950.00	950.00	950.00	950.00	950.00	950.00	950.00
H2S											
Cu2S											
Fe2O3					48.67	48.67	48.67		365.35		1096.06
H4Fe6(SO4)4(OH)14					8.16	8.16	8.16		61.25		183.85
H2SO4			100.00								
O2				100.00							
S											

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows PTA Flows (2)

Stream	PTAWH OF	PTA Wash	PTA Sol	PS Feed
Phase	L	L	SL	SL
No.	39	38	40	15
t/day	211.7	211.7	234.6	4287.5
kg/min	147.0	147.0	162.9	2977.4
l/min	125.9	147.0	89.6	2116.9
sg	1.168	1.000	1.818	1.406
	kg/min	kg/min	kg/min	kg/min
Cu++	4.90			164.20
Cu+				0.00
Zn++	0.30			10.40
Pb++				0.80
Fe++	4.60			165.93
Fe+++	1.80			61.17
Mg+	0.10			1.80
Cl-	15.50			525.10
SO4-				
HCl	-0.10			-2.00
H2O	119.60	147.00	85.20	2011.10
H2S				
Cu2S				
Fe2O3			66.60	33.30
H4Fe6(SO4)4(OH)14			11.16	5.58
H2SO4				
O2				
S				
	g/l	%	%	g/l
Cu++	38.62			77.56
Cu+	0.00			0.00
Zn++	2.44			4.89
Pb++	0.19			0.39
Fe++	36.60			78.38
Fe+++	14.40			28.90
Mg+	0.43			0.87
Cl-	123.51			248.06
SO4-				
HCl	-0.46			-0.93
H2O	950.00	100.00	50.00	950.00
H2S				
Cu2S				
Fe2O3			39.09	15.72
H4Fe6(SO4)4(OH)14			6.85	2.64
H2SO4				
O2				
S				

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows Pre-leach Flows (1)

Stream	Conc H2O	PL H2S	PL Feed	PL HCl	PL Prod	PLTH OF	PLTH UF	PLFL OF	PLFL UF	PL Recyc	CUP Out
Phase	L	G	SL	L	L	L	L	L	L	L	SL
No.	5	8	6	7	9	10	11	12	13	14	52
t/day	111.1	52.0	3022.9	250.2	3465.7	2599.3	866.4	649.8	216.6	1714.2	1308.7
kg/min	77.2	36.1	2099.2	173.7	2406.7	1805.0	601.7	451.3	150.4	1190.4	908.8
l/min	77.2		1516.7	147.2	1719.4	1289.6	429.9	322.4	107.5	850.4	666.3
sg	1.000		1.384	1.180	1.400	1.400	1.400	1.400	1.400	1.400	1.364
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
Cu++											
Cu+											
Zn++			26.8		45.2	33.9	11.3	8.5	2.8	22.4	4.4
Pb++			1.7		2.7	2.0	0.7	0.5	0.2	1.3	0.4
Fe++			225.2		268.4	201.3	67.1	50.3	16.8	132.8	92.5
Fe+++											
Mg*			7.0		11.6	8.6	3.0	2.2	0.8	6.2	0.8
Cl-			328.0		413.0	309.8	103.3	77.4	25.8	204.3	123.7
SO4-											
HCl			58.0	60.8	31.3	23.5	7.8	5.9	2.0	15.5	42.5
H2O	77.2		1440.9	112.9	1633.4	1225.1	408.4	306.3	102.1	807.9	632.9
H2S		36.1									
Cu2S			11.7								11.7
Fe2O3											
H4Fe6(SO4)4(OH)14											
H2SO4											
O2											
S											
	%	%	g/l	%	g/l	g/l	g/l	g/l	g/l	g/l	g/l
Cu++											
Cu+											
Zn++			17.67		26.31	26.31	26.31	26.31	26.31	26.31	6.65
Pb++			1.12		1.58	1.58	1.58	1.58	1.58	1.58	0.53
Fe++			148.50		156.12	156.12	156.12	156.12	156.12	156.12	138.77
Fe+++											
Mg*			4.61		6.70	6.70	6.70	6.70	6.70	6.70	1.18
Cl-			216.24		240.20	240.20	240.20	240.20	240.20	240.20	185.65
SO4-											
HCl			38.24	35.00	18.23	18.23	18.23	18.23	18.23	18.23	63.77
H2O	100.00		950.00	65.00	950.00	950.00	950.00	950.00	950.00	950.00	950.00
H2S		100.00									
Cu2S			7.71								17.56
Fe2O3											
H4Fe6(SO4)4(OH)14											
H2SO4											
O2											
S											

NOTES * - Mg = [Ca + Mg] expressed as Mg
** - MgO = [CaO + MgO] expressed as MgO

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows Pre-leach Flows (2)

Stream	OH Feed
Phase	L
No.	53
t/day	1534.9
kg/min	1065.9
l/min	761.5
sg	1.400
	kg/min
Cu++	
Cu+	
Zn++	20.0
Pb++	1.2
Fe++	118.9
Fe+++	
Mg+	4.6
Cl-	182.9
SO4-	
HCl	13.9
H2O	723.4
H2S	
Cu2S	
Fe2O3	
H4Fe6(SO4)4(OH)14	
H2SO4	
O2	
S	
	g/l
Cu++	
Cu+	
Zn++	26.31
Pb++	1.58
Fe++	156.12
Fe+++	
Mg+	6.00
Cl-	240.20
SO4-	
HCl	18.23
H2O	950.00
H2S	
Cu2S	
Fe2O3	
H4Fe6(SO4)4(OH)14	
H2SO4	
O2	
S	

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows FeCl3 Leach Flows (1)

Stream	PLFL UF	FS Feed	FS Prod	Preg	FSTH UF	Llx	SS Prod	SSTH OF	SSTH UF	SSFL OF	SSFL UF
Phase	L	SL	L	L	L	L	L	L	L	L	L
No.	13	15	16	17	18	19	20	21	22	23	24
t/day	216.6	4287.5	4659.8	3944.1	715.9	3100.7	4021.0	3509.6	511.4	383.6	127.9
kg/min	150.4	2977.4	3236.1	2738.9	497.1	2153.2	2792.4	2437.2	355.2	266.4	88.8
l/min	107.5	2116.9	2224.4	1882.7	341.7	1523.0	1877.3	1642.5	239.4	179.5	59.8
sg	1.400	1.406	1.455	1.455	1.455	1.414	1.484	1.484	1.484	1.484	1.484
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
Cu++		164.20	2.50		0.26	56.80	56.32	49.20	7.20	5.40	1.80
Cu+		0.00	242.20	207.10	37.34		107.88	94.20	13.70	10.30	3.40
Zn++	2.80	10.40	14.80	12.50	2.30	8.10	10.40	9.00	1.30	1.00	0.30
Pb++	0.20	0.80	1.20	1.00	0.20	0.60	0.80	0.70	0.10	0.10	0.00
Fe++	16.80	165.93	304.30	259.40	47.40	10.90	302.80	264.30	38.50	28.90	9.70
Fe+++		61.17	0.00	1.90	0.00	157.90	0.00	0.00	0.00	0.00	0.00
Mg+	0.80	1.80	2.60	2.20	0.40	1.40	1.80	1.60	0.20	0.20	0.10
Cl-	25.80	525.10	550.90	466.30	84.60	389.50	522.20	455.80	66.40	49.80	16.60
SO4-											
HCl	2.00	-2.00				49.40					
H2O	102.10	2011.10	2113.20	1788.60	324.60	1446.90	1783.40	1556.60	226.80	170.10	56.70
H2S											
Cu2S											
Fe2O3		33.30			33.30						
H4Fe6(SO4)4(OH)14		5.58			5.58						
H2SO4							2.30	2.00	0.30	0.23	0.07
O2											
S						31.60					
	g/l	g/l	g/l	g/l	g/l	g/l	g/l	g/l	g/l	g/l	g/l
Cu++		77.57	1.12	1.12	1.12	37.32	30.00	30.00	30.00	30.00	30.00
Cu+			108.88	108.88	108.88		57.47	57.47	57.47	57.47	57.47
Zn++	26.31	4.89	6.65	6.65	6.65	5.31	5.52	5.52	5.52	5.52	5.52
Pb++	1.58	0.39	0.53	0.53	0.53	0.42	0.44	0.44	0.44	0.44	0.44
Fe++	156.12	78.38	136.80	136.80	136.80	7.18	161.30	161.30	161.30	161.30	161.30
Fe+++		28.90	0.00	0.00	0.00	103.66	0.00	0.00	0.00	0.00	0.00
Mg+	7.30	0.87	1.18	1.18	1.18	0.94	0.98	0.98	0.98	0.98	0.98
Cl-	240.20	248.06	247.68	247.66	247.66	255.74	278.16	278.16	278.16	278.16	278.16
SO4-											
HCl	18.23	-0.93				32.44					
H2O	950.00	950.00	950.00	950.00	950.00	950.00	950.00	950.00	950.00	950.00	950.00
H2S											
Cu2S											
Fe2O3		15.72									
H4Fe6(SO4)4(OH)14		2.64									
H2SO4							1.21				1.21
O2											
S						20.78					

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows FeCl3 Leach Flows (2)

Stream	SS Wash	SSWH OP	Res Soln	PTA Feed
Phase	L	L	L	L
No.	25	26	27	28
t/day	574.5	361.4	341.0	4254.5
kg/min	398.9	250.9	236.8	2954.5
l/min	398.9	230.4		2048.0
sg	1.000	1.089	1.000	1.439
	kg/min	kg/min	kg/min	kg/min
Cu++		1.80		56.32
Cu+		3.40		107.88
Zn++		0.30		10.40
Pb++		0.00		0.80
Fe++		9.70		302.80
Fe+++		0.00		0.00
Mg+		0.10		1.80
Cl-		16.60		522.20
SO4-				
HCl				
H2O	398.90	218.90	236.80	1945.60
H2S				
Cu2S				
Fe2O3				
H4Fe6(SO4)4(OH)14				
H2SO4		0.07		2.30
O2				
S				
	%	g/l	%	g/l
Cu++		7.84		27.50
Cu+		14.82		52.67
Zn++		1.43		5.06
Pb++				0.40
Fe++		34.96		147.85
Fe+++		6.83		0.00
Mg+		0.25		0.90
Cl-		72.07		254.97
SO4-				
HCl				
H2O	100.00	950.00	100.00	950.00
H2S				
Cu2S				
Fe2O3				
H4Fe6(SO4)4(OH)14				
H2SO4		0.31		1.11
O2				
S				

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows — Electro-Win Flows

Stream	Preg	RW Out	Bleed	Anol In	Anol HCl	Anol Out	Str Out	CUP H2S	Cup Out
Phase	L	L	L	L	L	L	L	G	SL
No.	17	41	43	44	45	46	49	51	52
t/day	3944.1	3775.3	1336.0	2439.3	576.3	3012.5	1305.1	3.6	1308.7
kg/min	2738.9	2621.7	927.8	1694.0	400.4	2092.0	906.3	2.5	908.8
l/min	1882.7	1882.7	666.3	1216.4	339.3	1490.4	666.3		666.3
sg	1.455	1.393	1.393	1.404	1.180	1.404	1.360		1.364
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
Cu++						56.80			
Cu+	207.10	88.00	31.10	56.80		0.00	9.30		
Zn++	12.50	12.50	4.40	8.10		8.10	4.40		4.40
Pb++	1.00	1.00	0.40	0.60		0.60	0.40		0.40
Fe++	259.40	261.30	92.50	168.80		93.03	92.50		92.50
Fe+++	1.90					75.77			
Mg+	2.20	2.20	0.80	1.40		1.40	0.80		0.80
Cl-	466.30	398.60	141.10	257.60		337.40	128.90		123.70
SO4-									
HCl		69.60	24.60	45.00	140.10	103.00	37.10		42.50
H2O	1788.60	1788.60	632.90	1155.60	260.20	1415.80	632.90		632.90
H2S								2.50	
Cu2S									11.70
Fe2O3									
H4Fe6(SO4)4(OH)14									
H2SO4									
O2									
S									
	g/l	g/l	g/l	g/l	%	g/l	g/l	%	g/l
Cu++						38.14			
Cu+	110.00	46.72	46.71	46.71		0.00	14.02		
Zn++	6.65	6.65	6.65	6.65		5.43	6.65		6.65
Pb++	0.53	0.53	0.53	0.53		0.43	0.53		0.53
Fe++	137.77	138.77	138.77	138.77		62.41	138.77		138.77
Fe+++	1.00					50.84			
Mg+	1.18	1.18	1.18	1.18		0.96	1.18		1.18
Cl-	247.66	211.73	211.72	211.72		226.38	193.48		185.65
SO4-									
HCl		36.96	36.96	36.96	35.00	69.11	55.73		63.77
H2O	950.00	950.00	950.00	950.00	65.00	950.00	950.00		950.00
H2S								100.00	
Cu2S									17.56
Fe2O3									
H4Fe6(SO4)4(OH)14									
H2SO4									
O2									
S									

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows STA Flows

Stream	Anol Out	STA O2	STA H2S	Llx
Phase	L	G	G	SL
No.	46	47	48	19
t/day	3012.5	39.7	48.4	3100.7
kg/min	2092.0	27.6	33.6	2153.2
l/min	1490.4			1523.0
sg	1.404			1.414
	kg/min	kg/min	kg/min	kg/min
Cu++	56.80			56.80
Cu+	0.00			
Zn++	8.10			8.10
Pb++	0.60			0.60
Fe++	93.03			10.90
Fe+++	75.77			157.90
Mg+	1.40			1.40
Cl-	337.40			389.50
SO4-				
HCl	103.00			49.40
H2O	1415.80			1446.90
H2S			33.60	
Cu2S				
Fe2O3				
H4Fe6(SO4)4(OH)14				
H2SO4				
O2		27.60		
S				31.60
	g/l	%	%	g/l
Cu++	38.11			37.32
Cu+	0.00			
Zn++	5.43			5.31
Pb++	0.43			0.42
Fe++	62.41			7.18
Fe+++	50.84			103.66
Mg+	0.96			0.94
Cl-	226.38			255.74
SO4-				
HCl	69.11			32.44
H2O	950.00			950.00
H2S			100.00	
Cu2S				
Fe2O3				
H4Fe6(SO4)4(OH)14				
H2SO4				
O2		100.00		
S				20.78

TABLE VI Cont'd

GCM HYDROMETALLURGICAL COPPER PROCESS - Mass Balance
Detailed Stream Flows Oxyhydrolysis Flows

Stream	Evap Feed	Evap H2O	OH Feed	OH H2SO4	Evap O2	OH HCl	OH Sol	Fuel & Air Com Gas	
Phase	L	L	L	L	G	L	S	G	G
No.	53	54	55	56	57	58	59	57*	58*
t/day	1534.9	485.9	1049.0	62.9	4.32	831.0	347.9	803.9	1568.0
kg/min	1065.9	337.4	728.5	43.7	3.0	577.0	241.6	558.3	1088.9
l/min	761.5	337.4	424.1			489.1			
sg	1.400	1.000	1.718			1.180			
	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min	kg/min
Cu++									
Cu+								33.3	Oil
Zn++	20.0		20.0				20.0		
Pb++	1.2		1.2				1.2		
Fe++	118.9		97.5					404.0	404.0 N2
Fe+++			21.4						
Mg+	4.6		4.6				4.6		
Cl-	182.9		196.4				3.0		106.0 CO2
SO4-							42.8		
HCl	13.9					202.0			202.0 HCl
H2O	723.4	337.4	386.0			375.0			375.0 H2O
H2S									
Cu2S									
Fe2O3							170.0		
H4Fe6(SO4)4(OH)14									
H2SO4				43.7					
O2					3.0			121.0	1.9 O2
S									
	g/l	%	g/l	%	%	%	%	%	%
Cu++									
Cu+									
Zn++	26.31		47.10				8.34		
Pb++	1.58		2.83				0.50		
Fe++	156.12		230.00						
Fe+++			50.50						
Mg+	6.00		10.80				1.90		
Cl-	240.20		463.00				1.25		
SO4-							18.04		
HCl	18.23		0.00			35.00			18.55 HCl
H2O	950.00	100.00	910.00			65.00			34.44 H2O
H2S									
Cu2S									
Fe2O3							70.80		
H4Fe6(SO4)4(OH)14									
H2SO4				100.00					
O2					100.00				
S									

APPENDIX C

ENERGY BALANCE

TABLE VII

GCM Process Energy Requirement

PROCESS STAGE		BTU/LB CU NET	BTU/LB CU GROSS (1)	ENERGY SOURCE
Preleaching		557	696	steam
Leaching (gross)	Stage 1	600	700	steam
	Stage 2	-34		
Turboaeration (net)	PTA	-676		
	STA	-858		
Evaporation (gross)	2516			
Combustion Gases (net)	-926	56	70	steam
Turboaeration (net)	-1534			
Oxyhydrolysis		4337	4337	oil
TOTAL			5803	

	kWh/lb Cu
Electrowinning[2]	0.29
Other power	0.25
Total Electrical	0.54

1) Assumes 80% efficiency for steam generation and utilization.
2) Assumes a 10% distribution loss.

APPENDIX D

CAPITAL COSTS

TABLE VIII

GCM HYDROMETALLURGICAL COPPER PROCESS

Capital Cost Summary

AREA	15-Jun-87 COST
THAW SHED	$1,793,000
CONCENTRATE PREPARATION COMPLEX	$21,254,000
LEACHING PLANT	$27,591,000
PURIFICATION PLANT	$4,780,000
ELECTROWINNING & CASTING PLANT	$49,331,000
OXYGEN PLANT	$8,052,000
OXY-HYDROLYSIS PLANT	$30,597,000
SUBSTATION	$4,115,000
MISCELLANEOUS AREAS	$9,558,000
COPPER PROCESS PLANT TOTAL	$157,071,000

computer file:CCLEST01.WK1

TABLE IX

GCM HYDROMETALLURGICAL COPPER PROCESS

Capital Cost Detailed Estimate

15-Jun-87

AREA -Section	MATERIALS	LABOUR & CONTRACT & RENTAL	SPARES	OTHER (TAX & SHIP.)	ENGINEERING	PROJ. MAN.	CONTINGENCY	SECTION TOTAL
THAW SHED								
-Thaw Shed Building	356,000	894,000	0	34,000	188,000	22,000	299,000	1,793,000
CONCENTRATE PREPARATION COMPLEX								
-Conc. Prep. Building	2,534,000	4,172,000	30,000	238,000	668,000	115,000	1,551,000	9,308,000
-Concentrate Unloading	804,000	326,000	70,000	81,000	200,000	22,000	301,000	1,804,000
-Filter Area	2,815,000	847,000	152,000	271,000	320,000	66,000	894,000	5,365,000
-Concentrate Storage	687,000	421,000	31,000	67,000	200,000	21,000	285,000	1,712,000
-Feed Preparation	1,607,000	265,000	161,000	163,000	320,000	38,000	511,000	3,065,000
LEACHING PLANT								
-Leaching Plant Building	1,045,000	2,708,000	5,000	100,000	548,000	66,000	894,000	5,366,000
-Cu. Precip. and Preleach	2,881,000	771,000	171,000	101,000	560,000	67,000	910,000	5,461,000
-First Stage Leach	2,613,000	464,000	148,000	253,000	400,000	58,000	787,000	[illegible]
-Second Stage Leach	2,871,000	468,000	149,000	276,000	400,000	62,000	844,000	5,070,000
-Sulphur Filtration	841,000	387,000	73,000	86,000	360,000	26,000	355,000	2,128,000
-Primary Turboaeration	1,797,000	587,000	162,000	179,000	280,000	45,000	605,000	3,655,000
-Secondary Turboaeration	417,000	320,000	37,000	41,000	160,000	15,000	198,000	1,188,000
PURIFICATION PLANT								
-Purif. Plant Building	342,000	753,000	5,000	33,000	228,000	20,000	276,000	1,657,000
-Impurity Removal	1,260,000	554,000	101,000	128,000	520,000	38,000	522,000	3,123,000
ELECTROWINNING PLANT								
-Electro. Plant Building	2,831,000	5,176,000	83,000	273,000	1,002,000	140,000	1,901,000	11,406,000
-Copper Electrowinning	12,578,000	2,369,000	1,100,000	1,246,000	1,300,000	279,000	3,774,000	22,646,000
-Copper Billet Casting	1,392,000	8,973,000	622,000	197,000	1,360,000	188,000	2,547,000	15,279,000
OXYGEN PLANT								
-Site Preparation	61,000	70,000	0	6,000	94,000	3,000	47,000	281,000
-Turn-key Oxygen Plant	0	6,566,000	300,000	0	0	34,000	690,000	7,590,000
-Oxygen Pipeline	18,000	34,000	0	3,000	94,000	2,000	30,000	181,000
OXY-HYDROLYSIS PLANT								
-Site Preparation	999,000	733,000	0	93,000	336,000	32,000	439,000	2,632,000
-Turn-key Oxy-hyd. Package	0	22,900,000	1,000,000	0	0	120,000	3,603,000	27,623,000
-Oxy-hyd. Pipeline	86,000	92,000	0	9,000	94,000	4,000	57,000	342,000
SUBSTATION								
-20 MVA Substation	2,054,000	1,033,000	80,000	203,000	155,000	53,000	537,000	4,115,000
MISCELLANEOUS AREAS								
-Electrical	500,000	800,000	70,000	53,000	160,000	24,000	321,000	1,928,000
-Instrumentation	1,200,000	2,000,000	150,000	126,000	400,000	58,000	787,000	4,721,000
-Common Areas	958,000	1,036,000	60,000	94,000	240,000	36,000	485,000	2,909,000
TOTALS	45,547,000	65,719,000	4,760,000	4,354,000	10,587,000	1,654,000	24,450,000	157,071,000

computer file: CCLBST01.WK1

TABLE X

GCM HYDROMETALLURGICAL COPPER PROCESS

Capital Cost Comparison

15-Jun-87

Section	BD&A	HBM&S
Thaw shed	-	1,284,000
Feed preparation	5,200,000	2,196,000
Balance of feed prep. complex	2,775,000	13,546,000
Leaching plant	}	18,564,000
Purification plant	} 41,253,000	3,176,000
Electrowinning plant	}	25,656,000
Billet Casting	-	11,184,000
Oxygen plant	7,200,000	7,058,000
Sulphur filtration	6,336,000	1,387,000
Oxy-hydrolysis plant	19,200,000	25,912,000
Substation	3,050,000	3,370,000
Tailings pond	200,000	-
Miscellaneous	-	7,047,000
DIRECT PLANT COSTS	85,214,000	120,380,000
Engineering	13,208,000	10,587,000
Proj. man. (contractor's fee)	2,201,000	1,654,000
Contingency	13,208,000	24,450,000
TOTAL	$113,831,000	$157,071,000

computer file:CCLEST01.WK1

TABLE XI

GCM Hydrometallurgical Copper Process

Mechanical Equipment List

QTY.	DESCRIPTION	SIZE	MOTORS
	CONCENTRATE PREPARATION COMPLEX		
	CONCENTRATE PREPARATION BUILDING		
1	Bridge crane	5 t - 15 m span	2 drives
	CONCENTRATE UNLOADING		
1	Railcar tugger	12 car capacity	20 HP
1	Joslin articulated unloader		150 HP
2	Railcar cap hoists	5 t	1 each
2	Slot feeders	1200 mm x 10 m	1 each
1	Conveyor from Joslin to storage	1000 mm x 52 m	1 drive
1	Conveyor from Joslin to storage	1000 mm x 22 m	1 drive
2	Conc. receiving hoppers	70 t	
	FILTER AREA		
4	Conc. storage tank (mild steel)	5 m dia x 6 m	
3	Pressure filter feed and filtrate tanks (mild steel)	5 m dia x 6 m	
6	Tank agitators		20 HP
3	Disc filters for Flin Flon conc. (moved from existing filter plant)		2 drives
2	Pressure filters for dewatering chloride leach feed	50 chambers	6 drives
2	Manual hoist above pressure filters	10 t	
1	Dust collection scrubber		1 drive
1	Conveyor under disc filters	1000 mm x 21 m	1 drive
2	Conveyors under pressure filters	1000 mm x 10 m	1 each
1	Conveyor to transfer point	1000 mm x 44 m	1 drive
	CONCENTRATE STORAGE		
4	Slot feeders	1200 mm x 5 m	1 each
1	Conveyor from railcar unloading	1000 mm x 90 m	1 drive
1	Conveyor to grinding	1000 mm x 120 m	1 drive
4	Hoppers in conc. storage	15 t	
	FEED PREPARATION		
2	Vibrating feeders	40 tph	1 each
2	Ball mills		600 HP
2	Cyclone packs		
1	Grinding feed hopper	100 t	

TABLE XI cont'd

GCM Hydrometallurgical Copper Process

Mechanical Equipment List

QTY.	DESCRIPTION	SIZE	MOTORS
	LEACHING PLANT		
	LEACHING PLANT BUILDING		
1	Bridge crane	5 t - 15 m span	2 drives
	Cu PRECIP. AND PRELEACH		
3	Cu precip. tanks (FRP)	3.2 m dia x 5 m	
4	Preleach tanks (FRP)	5.3 m dia x 8 m	
2	Pressure filter surge tanks (FRP)	3.5 m dia x 4 m	
8	Tank agitators		30 HP
2	Pressure filter for dewatering preleach solution	50 chambers	6 drives
2	Conveyors under pressure filters	1000 mm x 10 m	1 each
2	Conveyors feeding first stage leach	1000 mm x 15 m	1 each
2	Ventilation fans		20 HP
	FIRST STAGE LEACH		
4	First stage leach tanks (FRP)	3.5 m dia x 4 m	
2	Pressure filter surge tanks (FRP)	3.5 m dia x 4 m	
5	Tank agitators		30 HP
2	Pressure filter for dewatering first stage leach solution	50 chambers	6 drives
2	Conveyors under pressure filters	1000 mm x 10 m	1 each
2	Conveyors feeding first stage leach	1000 mm x 15 m	1 each
	SECOND STAGE LEACH		
4	First stage leach tanks (FRP)	3.5 m dia x 4 m	
2	Pressure filter surge tanks (FRP)	3.5 m dia x 4 m	
5	Tank agitators		30 HP
2	Pressure filter for dewatering second stage leach solution	60 chambers	6 drives
2	Conveyors under pressure filters	1000 mm x 12 m	1 each
2	Conveyors feeding first stage leach	1000 mm x 15 m	1 each
	SULPHUR FILTRATION		
1	Conveyor feeding melting cyclone	1000 mm x 10 m	1 drive
1	Sulphur melting cyclone		
2	"Sparkler" sulphur filters	250 tpd	2 drives
2	Ventilation units		1 each
1	Molten sulphur transport		

TABLE XI cont'd

GCM Hydrometallurgical Copper Process

Mechanical Equipment List

QTY.	DESCRIPTION	SIZE	MOTORS
	LEACHING PLANT CONT'D		
	PRIMARY TURBOAERATION		
4	Turboaeration tanks (FRP)	6 m dia x 8 m	
2	Pressure filter surge tanks (FRP)	3.5 m dia x 4 m	
4	Tank agitators		75 HP
1	Tank agitator		30 HP
1	Pressure filter for dewatering turboaeration solution	50 chambers	6 drives
1	Reversible conveyor under pressure filter	1000 mm x 10 m	1 drive
1	Blower fan		20 HP
	SECONDARY TURBOAERATION		
4	Turboaeration tanks (FRP)	5.8 m dia x 6 m	
4	Tank agitators		50 HP
2	Blower fans		20 HP
	PURIFICATION PLANT		
	PURIFICATION PLANT BUILDING		
1	Bridge crane	5 t - 18 m span	2 drives
	IMPURITY REMOVAL		
3	Settling cone O/F tanks (FRP)	3.5 m dia x 4 m	
1	Silver removal tank (FRP)	4.6 m dia x 6 m	
2	Table filter feed tanks (FRP)	3.5 m dia x 4 m	
3	Tank agitators		20 HP
3	Settling cones	4 m dia	
2	Table filters	1 square metre	2 drives
2	Filter presses	20 chambers	2 drives
2	Screw conveyors under filters		1 drive
2	Ventilation units		20 HP
1	Column reactor		
	ELECTROWINNING PLANT		
	ELECTROWINNING PLANT BUILDING		
1	Bridge crane	5 t - 18 m span	2 drives
1	Bridge crane	30 t - 20 m span	2 drives

TABLE XI cont'd

GCM Hydrometallurgical Copper Process

Mechanical Equipment List

QTY.	DESCRIPTION	SIZE	MOTORS
	ELECTROWINNING PLANT CONT'D		
	COPPER ELECTROWINNING		
2	Solution storage tanks (FRP)	6 m dia x 8 m	
3	Cell feed tanks (FRP)	4 m dia x 6 m	
3	Settling cone O/F tanks (FRP)	3 m dia x 5 m	
1	Mixing tank (FRP)	3 m dia x 5 m	
1	Tank agitator		30 HP
4	Heat exchangers		
3	Settling cones	4 m dia	2 drives
2	Primary copper table filter	15 square metres	2 drives
2	Stripper copper table filter	5 square metres	2 drives
10	Ventilation units		20 HP
12	Primary electrowinning sections	60 cathodes each	
4	Secondary electrowinning sections	60 cathodes each	
	COPPER BILLET CASTING		
2	Copper cake conveyors to storage	500 mm x 15 m	1 drive
2	Slot feeders below storage	500 mm x 5 m	1 drive
2	Screw conveyors feeding dryers		1 drive
2	Fluid bed dryers	10 tph each	3 drives
2	Conveyors feeding forming equipment	500 mm x 15 m	1 drive
2	Roll type briquette forming equip.		1 drive
1	Screw conveyor feeding furnace		1 drive
1	Slot feeder below flux unloading	1000 mm x 4 m	1 drive
1	Flexible wall conveyor	approx. 30 m long	1 drive
1	Vibrating screen for coarse flux		1 drive
2	Flux conveyors feeding dryers		1 drive
1	Shaft type melting furnace and off-gas handling equipment		
1	Holding furnace and billet casting facility		
1	Rubber tired ladel transport		
	OXY-HYDROLYSIS PLANT		
	SITE PREPARATION		
2	Oil storage tanks	10 m dia x 10 m	
1	Railcar oil unloading facility		
1	Oil pumping facility		
	MISCELLANEOUS AREAS		
	COMMON AREAS		
1	Conveyor from feed prep. to leaching	1000 mm x 50 m	1 drive
3	Air compressors for filters	1500 cfm	300 hp

TABLE XII

GCM Hydrometallurgical Copper Process

Process Flow Pump List

QTY.	DESCRIPTION
	CONCENTRATE PREPARATION COMPLEX
	CONCENTRATE UNLOADING
*	Hydraulic pumps for Joslin unloader
	FILTER AREA
2	Disc filter feed pump
2	Pressure filter feed pump
2	Pressure filter wash water pump
	FEED PREPARATION
2	Ball mill discharge pump (to cyclone)
	LEACHING PLANT
	Cu. PRICIP. AND PRELEACH
2	Preleach discharge pump
2	Pressure filter feed pump
2	Pressure filter wash water pump
2	Pressure filter filtrate pump
	FIRST STAGE LEACH
2	First stage discharge pump
2	Pressure filter feed pump
2	Pressure filter wash water pump
2	Pressure filter filtrate pump
	SECOND STAGE LEACH
2	Second stage discharge pump
2	Pressure filter feed pump
2	Pressure filter wash water pump
2	Pressure filter filtrate pump
	SULPHUR FILTRATION
2	"Dirty" molten sulphur recycling pump
2	"Sparkler" filter feed pump
2	Clean molten sulphur pump

GCM Hydrometallurgical Copper Process

Process Flow Pump List

QTY.	DESCRIPTION
	LEACHING PLANT CONT'D
	PRIMARY TURBOAERATION
2	Primary turboaeration discharge pumps
2	Pressure filter feed pump
2	Pressure filter wash water pump
2	Pressure filter filtrate pump
	SECONDARY TURBOAERATION
2	Secondary turboaeration discharge pumps
	PURIFICATION PLANT
	IMPURITY REMOVAL
2	Column reactor discharge pump
2	Gold settling thickener O/F puimp
2	Gold settling thickener U/F puimp
2	Silver settling thickener O/F puimp
2	Silver settling thickener U/F puimp
2	First stage silver removal pump
2	First stage table filter feed pump
2	Second stage table filter feed pump
2	Repulped first stage table filter cake pump
	ELECTROWINNING PLANT
	COPPER ELECTROWINNING
2	Catholyte solution storage discharge pump
6	Catholyte feed pumps to primary cells
6	Catholyte feed pumps to stripper cells
6	Anolyte feed pumps to prim. and strip. cells
2	Feed pump to stripper cell catholyte tank
2	Catholyte feed tank pumps to copper precip.
2	Anolyte feed tank pumps to anolyte sol. stor.
2	Copper settling cone overflow pumps
2	Copper settling cone underflow pumps
2	Stripped copper mixing tank discharge pumps
	OXYGEN PLANT
	-Turn key plant-
	OXY-HYDROLYSIS PLANT
	-Turn key plant-

APPENDIX E

OPERATING COSTS

TABLE XIII

GCM HYDROMETALLURGICAL COPPER PROCESS
OPERATING COSTS

	B.D.& A.	REVISED by H.B.M.& S.
DIRECT COSTS	cents/lb COPPER	cents/lb COPPER
REAGENTS		
grinding media	0.20	0.17
HCl	1.32	1.18
H2SO4	0.93 *	0.70
	2.45	2.05
UTILITIES		
electrowinning power	0.66 *	0.66
other power	0.57	0.57
process steam		0.87
oxy-hydrolysis fuel	\	2.59
dryer oil	2.57	0.09
melter oil	/	0.10
oxygen		0.36
water	0.10	0.10
	3.90	5.34
DIRECT LABOUR		
labour	3.90	2.54
supervision	0.78	0.30
fringe benefits		0.80
	4.68	3.64
MAINTENANCE		
labour	1.44	1.06
materials	0.74	1.12
supervision and overhead	0.36	1.12
	2.54	3.30
PAYROLL OVERHEAD		
35% of labour	2.27	
TOTAL DIRECT COSTS	15.84	14.33
INDIRECT COSTS		
overhead		1.95
40% of direct lab. and maint.	2.89	
TOTAL INDIRECT COSTS	2.89	1.95
FIXED COSTS		
taxes	0.74	0.74
insurance	0.74	0.74
TOTAL FIXED COSTS	1.48	1.48
TOTAL OPERATING COSTS	20.21 cents/lb.	17.76 cents/lb.

* Corrected figures from B.D.& A. report

computer file: CCLOPCST.WK1

TABLE XIV

BASIS FOR OPERATING COSTS

1. Grinding Media

The cost of grinding media for concentrate regrind was taken as 30 cents per pound steel consumed.

2. Hydrochloric Acid

The hydrochloric acid required is based on a loss of 3% of the HCl equivalent of the total chloride going to the oxyhydrolysis circuit. An acid cost of \$125/t (20°) Baume was used, plus \$92/t for freight to Flin Flon.

HCl required = 0.0430 kg/kg Cu

3. Sulfuric Acid

Sulfuric Acid cost is estimated at \$50/t based on burning the by-product S° produced.

H_2SO_4 required = 0.342 kg/kg Cu

4. Power

Power costs are based on 2.3¢/kwh. Electrowinning power required is 0.63 kwh/kg Cu which includes a 10% distribution loss. Other power required is estimated at 0.55 kwh/kg Cu.

5. Fuel Oil

Fuel oil costs are based on 23.8¢/l. Fuel oil required is estimated at 0.34 l/kg Cu (0.24 l/kg Cu for oxyhydrolysis).

6. Oxygen

Oxygen is provided from the ZPL oxygen plant. Incremental oxygen costs are estimated at \$18.50/tonne.

Oxygen required is 0.43 kg/kg Cu.

7. Labour

Direct labour is based on \$28,000/y. Fringe benefits are 30%.

8. Maintenance

Costs are based on similar equipment maintenance costs at HBM&S.

APPENDIX F

FINANCIAL EVALUATION AND CRITERIA

TABLE XV

FINANCIAL ANALYSIS SUMMARY

GCM HYDROMETALLURGICAL COPPER PROCESS

(CONSTANT 1986 $)

YEAR	1	2	3	4	10
Change in Revenues From Products					
Copper				3,547,416	3,547,416
Zinc				9,584,442	9,584,442
Sulphur				7,554,326	7,554,326
Gold				(14,000,017)	(14,000,017)
Silver				(462,503)	(462,503)
Total Revenue	0	0	0	6,223,664	6,223,664
Marketing & Shipping	0	0	0	(4,527,391)	(4,527,391)
Total Revenue Change	0	0	0	1,696,273	1,696,273
Operating Costs					
Cu Extraction				3,206,789	3,206,789
Zn extraction				4,586,973	4,586,973
Cu & P.M. Refining				(13,437,299)	(13,437,299)
Total Operating Costs	0	0	0	(5,643,537)	(5,643,537)
TOTAL NET BENEFIT	**0**	**0**	**0**	**7,339,810**	**7,339,810**
Capital costs					
Chloride leach	45,000,000	70,000,000	42,071,000		
Total capital costs	45,000,000	70,000,000	42,071,000	0	0
PRETAX CASH FLOW	(45,000,000)	(70,000,000)	(42,071,000)	7,339,810	7,339,810
CASH FLOW TO DATE	(45,000,000)	(115,000,000)	(157,071,000)	(149,731,190)	(83,672,900)

GCM financial analysis
1986 GCM (assumes ZPL commissioned)
computer file: GCMIRR.WK1 10-Jun-87

TABLE XVI

FINANCIAL ANALYSIS CRITERIA

GCM HYDROMETALLURGICAL COPPER PROCESS METAL DISTRIBUTION

COPPER CONCENTRATE

COPPER - copper is extracted in two stages of leaching and after impurity removal is electrowon. Stripper copper product will require further refining because of impurities present.

98.0% extraction	x 84.5%	reporting to copper product
	x 15.5%	to stripper copper product
2.0%	x 100.0%	reporting to 2nd stage leach residue

ZINC - zinc is extracted in the pre-leach and main leach stages and recovered in oxyhydrolysis as zinc sulphate. The menatite (containing zinc sulphate) can be stockpiled with zinc recovered through run-off.

98.0% extraction	x 85.5%	recovery from stockpile to zinc hydroxide or zinc sulphate
2.0%	x 100.0%	reporting to 2nd stage leach residue

GOLD - although some gold is extracted in the leach (BDA report p. 99) and can be recovered as a high grade product, the major portion remains in the leach residue. For the purpose of this study, it is assumed that all of the gold ends up in leach residue.

SILVER - silver extraction varies (BDA report page 99) ranging from 93% to 96%. An average extraction of 95% is assumed.

95.0% extraction	x 99.0%	reporting to silver product
	x 1.0%	reporting to copper product
5.0%	x 100.0%	reporting to 2nd stage leach residue

TABLE XVI –Cont'd

NOTE: Silver product will be combined with stripper copper for further treatment at a refinery.

SUMMARY Recoveries are presumed the same for HUDBAY and purchased conc.

copper conc. – product copper	82.8%
copper conc. – stripper copper	15.2%
copper conc. – zinc to ZPL	83.8%
copper conc. – gold to GCM residue	100.0%
copper conc. – silver product	94.0%
copper conc. – silver in copper	1.0%
copper conc. – silver in GCM residues	5.0%

TABLE XVII

FINANCIAL ANALYSIS CRITERIA

INCREMENTAL OUTPUTS & REVENUES

Products	Base Case (ZPL & Cu Smelter	GCM Option (GCM & ZPL)	Incremental Change
Copper (tonnes/y)			
electrorefined	71 462	11 367	
electrowon	-	61 966	
TOTAL CU	71 462	73 333	+1 871
Zinc (tonnes/y)	82 359	90 360	+8 001
Sulfur (tonnes/y)	36 590	92 490	+55 900
Silver (oz/y)			
in Cu annodes	1 138 899	-	
in Ag product	-	1 055 620	-41 761
in ZPL residue	-	41 518	
Gold (oz/y)			
in Cu anodes	81 946	-	+2 797
in residues	-	84 743	

Revenues	$C/Y		
Copper	135 491 952	139 039 368	+3 547 416
Zinc	98 007 210	107 591 652	+9 584 442
Sulfur	4 944 773	12 499 099	+7 554 326
Silver			
in Cu anodes	8 462 020	-	
in Ag product	-	7 843 257	-462 503
in ZPL residue	-	156 260	
Gold			
in Cu anodes	45 402 181	-	
in residues	-	31 402 164	-14 000 017

TABLE XVIII

FINANCIAL ANALYSIS CRITERIA

BASIS FOR REVENUES

Prices	$U.S.	$CDN.
$C = 0.74 $US		
Copper	-	1.896/kg
Zinc	-	1.190/kg
Gold	410	554.05/oz
Silver	5.50	7.43/oz
Sulfur	100	135.14/tonne

TREATMENT CHARGES:

GCM Process Residue:

FeS_2 = 45%
S^o = 29%
Gangue = 24%
Cu = 0.6%

Au = 0.951 oz/tonne
Ag = 0.687 Oz/tonne

There is no payment for copper and silver

	$C/tonne Residue	
Gross Revenue	$526.90	
Processing Charge	75.00	
Recovery Penalty (5%)	26.34	172.10
Refining Charge ($5/oz)	4.76	
Shipping Charge	66.00	
Net Revenue	$354.80	

Annual Production = 81 713 tonnes
Annual Revenue = $28 991 722

TABLE XVIII - Cont'd

ZPL Process Residue:

FeS_2 = 45%
S^o = 29%
Zn = 13%
Cu = 2%
Gangue = 11%

Au = 0.74 oz/tonne
Ag = 4.364 oz/tonne

There is no payment for copper or zinc.
There is a silver deduction of 0.55 oz/tonne

	$C/tonne Residue			
	Au	Ag	Total	
Gross Revenue	410.00	28.34	438.34	
Processing Charge			75.00	
Recovery Penalty (5%)	20.50	1.42	21.92	
Refining Charge				
$5/oz Au	3.70	-	5.61	168.53
$0.50/oz Ag	-	1.91		
Shipping Charge			66.00	
NET REVENUE	(385.80)	(25.01)	269.81	

Annual Production = 9 513 tonnes
Annual Revenue = $2 566 702 [approx. 94% from gold]
[approx. 6% from silver]

TOTAL ANNUAL REVENUE FROM GCM & ZPL RESIDUES = $31 558 424

TABLE XIX

FINANCIAL ANALYSIS CRITERIA

INCREMENTAL CHANGES IN OPERATING COSTS

FOR GCM OPTION

Operation	$C/Y	Rationale
Cu Extraction	+ 3 206 789	GCM process versus Base Case
Cu Refining	-12 985 929	Only stripper copper (11,367 tonnes/y) requires refining.
Zn Extraction	+ 4 586 973	Total costs for recovering saleable zinc from the copper concentrate are estimated at 23¢/lb zinc. However, because of the uncertainty related to chloride contamination a contingency of 3¢/lb was added, so that total credit from zinc recovery essentially equals the cost of replacement concentrate.
PM Refining		No gold is recovered from
Au	- 409 730	Cu anodes, and silver
Ag	- 41 640	output (with Cu) is 83 279 oz/y less than base case.
Total Incremental Operating Costs	- 5 643 537	
Sulphur Shipping	+ 3 689 400	Based on $66.00/tonne for annual production of 55 900 tonnes.
Copper Marketing & Shipping	+ 185 229	Based on increased copper output at $99/tonne.
Zinc Marketing & Shipping	+ 652 762	Based on increased zinc output at $81.60/tonne.
TOTAL INCREMENTAL MARKETING/SHIPPING COSTS	+ 4 527 391	